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DESIGN GUIDES FOR BIOLOGICAL WASTEWATER TREATMENT PROCESSES

Joseph F. Malina, et al

Texas University
Austin, Texas

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Design Guides for Biological Wastewater Treatment Processes

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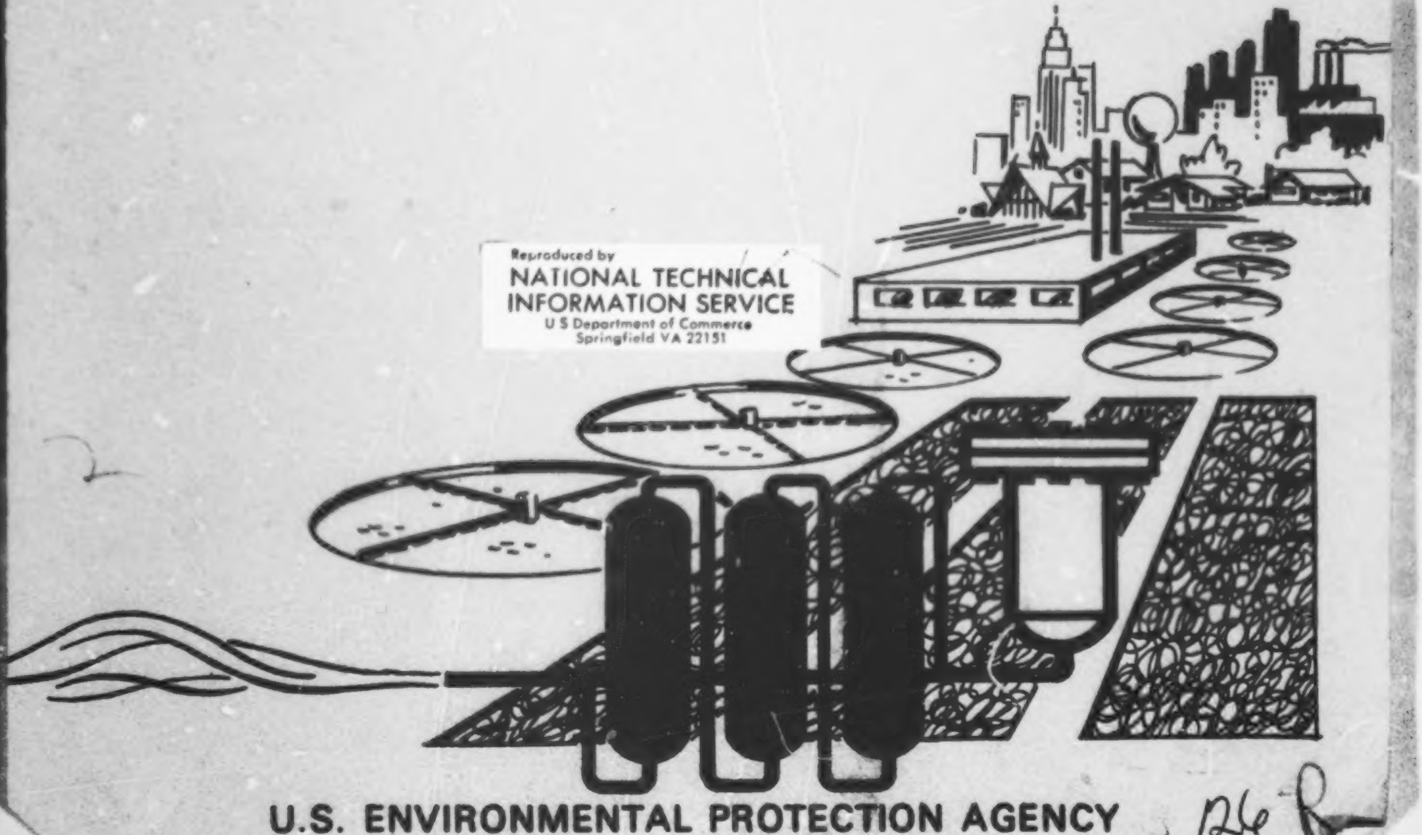

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Design Examples

27 Abstract

This report provides a set of guidelines for the design of biological processes for the treatment of municipal wastewater. The equations and factors which must be considered in the design of the activated sludge system, the contact stabilization system, trickling filter plants, aerated lagoons, and waste stabilization ponds are identified. The applicability and limitations of each system and mathematical model of each process are established. Operating data from the Govalle Wastewater Treatment Plant, the Williamson Treatment Plant, and the Walnut Creek Plant operated by the City of Austin, Texas and other operating data from the treatment plants where sufficient applicable data were recorded were used to develop rate constants and other coefficients required for application of the mathematical models and other design of treatment plants. The need for waste characterization including variations in quantities in flow and composition of flow are emphasized. The significant design considerations are discussed, design procedures are outlined and design calculations are developed. This report contains 211 pages, 81 figures, ten tables, and 88 references.

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by

THE CITY OF AUSTIN, TEXAS

and

CENTER FOR RESEARCH IN WATER RESOURCES
Environmental Health Engineering Research Laboratory
Civil Engineering Department
The University of Texas
Austin, Texas

for the

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ABSTRACT

The objective of this report is to provide a set of guidelines for the design of various biological treatment processes. The equations and factors which must be considered in the design of the activated sludge system, the contact stabilization system, trickling filter plants, aerated lagoons, and waste stabilization ponds were based on operating data from full-scale plants at the Govalle Wastewater Treatment Plant, the Williamson Treatment Plant, and the Walnut Creek Plant operated by the City of Austin, Texas and other operating data from the treatment plants where sufficient applicable data were recorded.

The need for waste characterization including variations in quantities in flow and composition of flow are emphasized. The applicability and limitations of the design equations are presented. The significant design considerations are discussed and design calculations included where these calculations would be meaningful and in other cases, a design procedure is outlined in detail.

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CONCLUSIONS

The conclusions presented below are based on the results of laboratory and field-scale experimentation with biological treatment processes at the Govalle Wastewater Treatment Plant, the Williamson Creek Wastewater Treatment Plant, and the Walnut Creek Wastewater Treatment Plant, all operated by the City of Austin, Texas.

1. Wastewater in Austin, Texas is from domestic and residential origin and the industrial contributions are insignificant in volume and concentration. The rate of flow and the strength of the wastewater are markedly influenced by the infiltration of rainwater into the collection system. The average concentrations of the influent BOD and suspended solids is 155 mg/l. The average flow of wastewater to the Govalle Plant on Sundays is 20 MGD and on work days is 23.5 MGD. An average dry weather flow of 25 MGD will occur 90 percent of the time whereas a flow of 30 MGD will occur 90 percent of the time during periods of rainfall.

2. The conventional activated sludge process resulted in effluent suspended solids concentrations of 12 to 22 mg/l, a total BOD of 20 to 23 mg/l, and a soluble BOD of three to seven mg/l. The activated sludge process was evaluated at loading rates of 0.23 to 0.28 pounds of BOD per pound of mixed liquor suspended solids per day resulting in BOD removal rates of 0.198 to 0.234 lb BOD/lb MLSS - day. The narrow range of organic loadings resulted from dilution of the incoming wastewater during periods of heavy rainfall. The system was operated at aeration times which ranged from 2.2 to 4.9 hours.

3. The results of laboratory scale and field-scale evaluation of the contact stabilization process indicate that effluent concentrations of soluble BOD of six to 11 mg/l are possible at contact times as low as 15 minutes. In systems in which the contact and stabilization basins are not physically separated, back-mixing of the wastewater occurs at the point of introduction into the aeration basin. Therefore, the contact time is somewhat longer than the theoretical value. The data indicate that the average reduction in BOD for the contact stabilization system is 78 percent. BOD loading rates of 0.2 to 0.5 lb BOD/lb MLSS - day based on the total quantity solids in the aeration system were applied to the contact stabilization process.

4. Results of the studies of the pilot-scale trickling filter with corrugated plastic medium indicate that the soluble effluent BOD of 9.5 and 27.5 mg/l are obtainable at hydraulic loadings of 63 and 189 MGAD, respectively. These hydraulic loadings resulted in organic loading rates of 38 and 161 lb BOD/1000 cu ft-day. The performance of the trickling filter process is controlled to a large extent by the type of medium used. The hydraulic characteristics of the medium affects the magnitude of the overall BOD removal rate constant.

5. The data observed for the field-scale aerated lagoons indicate that soluble effluent BOD concentrations of less than five mg/l are possible at detention times of about 2.5 days. At detention times of less than one day, the effluent soluble BOD was between seven and 19 mg/l. The performance of the aerated lagoons are affected by the operating temperature, the power level, and the concentration of suspended solids in the aeration basin. The system is less efficient at low temperatures than at high temperatures. A power level of 30 horsepower per million gallons or higher results in a completely mixed system. At power levels of less than 30 horsepower per million gallons, some of the suspended solids settle and a zone of anaerobic activity will exist at the bottom of the lagoon. An aerated lagoon must be followed by an additional pond or clarifier to reduce the effluent suspended solids concentration to a level prescribed by most regulatory agencies. Removal of suspended solids will also markedly reduce the total BOD of the effluent.

6. The results of these investigations of stabilizationponds indicate that a three pond system consisting of a separate anaerobic pond with a short detention time followed by a facultative pond and a smaller maturation pond produced the best effluent of the three systems evaluated. The effluent quality expressed as total BOD increased as the areal loading increased from 47 to 165 lb BOD_u/acre-day. However, the soluble effluent BOD was independent of loading rate at the range of loading used in this study, and the soluble BOD concentration was between two and ten mg/l. The soluble effluent BOD concentration was also in the same range for the other two systems.

7. Any of the biological systems described above can be used for the effective treatment of municipal wastewaters provided the proper design equation is applied and the proper design considerations are included in developing the system.

RECOMMENDATIONS

Design of municipal wastewater treatment plants should be based on the best available information and should take into consideration the variations in flow and composition of the wastewater as well as the average flow and quality characteristics. In those communities where a collection system is already installed reliable quantity and composition data can be developed by a survey of the existing system. In those communities where a collection system does not exist, published information can be used as a basis for design provided the information selected from the published sources is for communities which have about the same population, commercial and industrial activities, and climate.

The design equations presented in this report can be used for the design of the unit processes selected for the biological treatment of municipal wastewater. The coefficients and the rate constants which are included in the discussions can be used for preliminary design, provided no other information is available.

Design of the biological treatment processes should take into consideration the type of preliminary or primary treatment employed. The type of sludge handling and disposal system provided may have a bearing on the selection of the biological treatment system. The available land and the degree of treatment required by effluent standards may dictate the type of system required and eliminate the use of other biological treatment processes.

The performance of a wastewater treatment facility will be only as good as the operation and maintenance of the system. Therefore, the qualifications of the operator should be taken into consideration during the design of the particular system.

MUNICIPAL WASTEWATER CHARACTERISTICS

INTRODUCTION

The design of municipal wastewater treatment facilities must take into consideration the quantity and composition of the wastewater. The average and the range of flow must be included in the hydraulic design of the treatment plant. The characteristics of the wastewater will dictate the type of unit processes necessary for effective treatment.

The quantity and quality of the influent wastewater is affected by the land use of the drainage area, the extent to which sanitary and storm water are separated, the amount of infiltration, the rainfall pattern and the type of industrial waste ordinance enforced by the municipality. The wastewater generated in a particular area of a city is related to the water use pattern which in turn is established by the price of water and the type of development of the land. For example, the water use pattern is different for single family residences, apartments, commercial and industrial developments. Industries which operate only seasonally or which have batch processes and institutions which have large transient populations such as universities can markedly affect the quantity of wastewater which must be treated.

WASTEWATER FLOW

The necessary data can be collected for particular cities if the wastewater collection system is already installed. The hydraulic characteristics of the system and variations of the wastewater flows from different parts of the city can be determined by field surveys. The quality characteristics of the wastewater from the various areas can also be determined as part of the field survey. The treatment plant can be designed based on the existing land use pattern and the respective qualitative and quantitative characteristics of the return flows. The projection of future land use patterns and the anticipated population densities will permit accurate estimates of the projected design flow to the proposed treatment facility. The quantity of wastewater return flow generated in different portions of typical cities are available in many of the classical textbooks or design manuals. (ASCE, 1960; Billings, 1971; Clark, 1971; Fair, 1966; Fair, 1971; Steel, 1960) However, many of the reported values are for municipalities which do not correspond completely to the climatological and land use pattern of the municipality for which the treatment plant is designed. Therefore, in those municipalities where the water distribution and wastewater collection systems exist, it is a relatively easy task to correlate the water consumption and return flows for various areas of the municipality. In this manner a more accurate estimate of the return flow can be derived. The quality characteristics of the composite wastewater can also be developed from these data.

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In those cities where a wastewater collection system does not exist, it is necessary to estimate the quantity and quality of wastewater. The land use pattern and zoning regulations must be established so that an accurate estimate of the population density can be developed for the present population and for the projected future population. The total population to be served by the municipal treatment plant must also be determined in order to develop the flow for the design of the facility. The estimated water use and wastewater generation patterns must be compared with and verified by the results of other surveys conducted in communities of similar size and the same geographical and climatological area. The industrial contribution to the total flow and the characteristics of these return flows must also be carefully defined.

The total flow generated by a community and the composition of the wastewater are a function of the population served and the industrial development in the particular municipality. The data in Figure 1-1 indicate the relationship between population served and the average wastewater flow for communities of populations between 2,000 and 50,000 in the State of Texas (Williamson, 1970). The range of average wastewater flows for communities of the same population is wide. A statistical evaluation of the available data can be used to determine the design flow. Graphical analyses of return flow data are presented in Figure 1-2 (Williamson, 1970). The per capita flow increases as the population served increases. These data are for municipal flow with very little or no industrial flow included in these data. The composition of the wastewater is also affected by the population of the municipality and in turn by the quantity of water used. The flow data exhibited a geometrically normal distribution and the statistical parameters describing the wastewater flow data are summarized in Table 1-1.

Table 1-1

Wastewater Flows (Williamson, 1970)

Population	Geometric Mean Flow (gallons/capita-day)	Geometric Standard Deviation
2,500-4,999	67	1.47
5,000-9,999	74	1.61
10,000-14,999	75	1.62
15,000-24,999	89	1.64
25,000-49,999	101	1.36

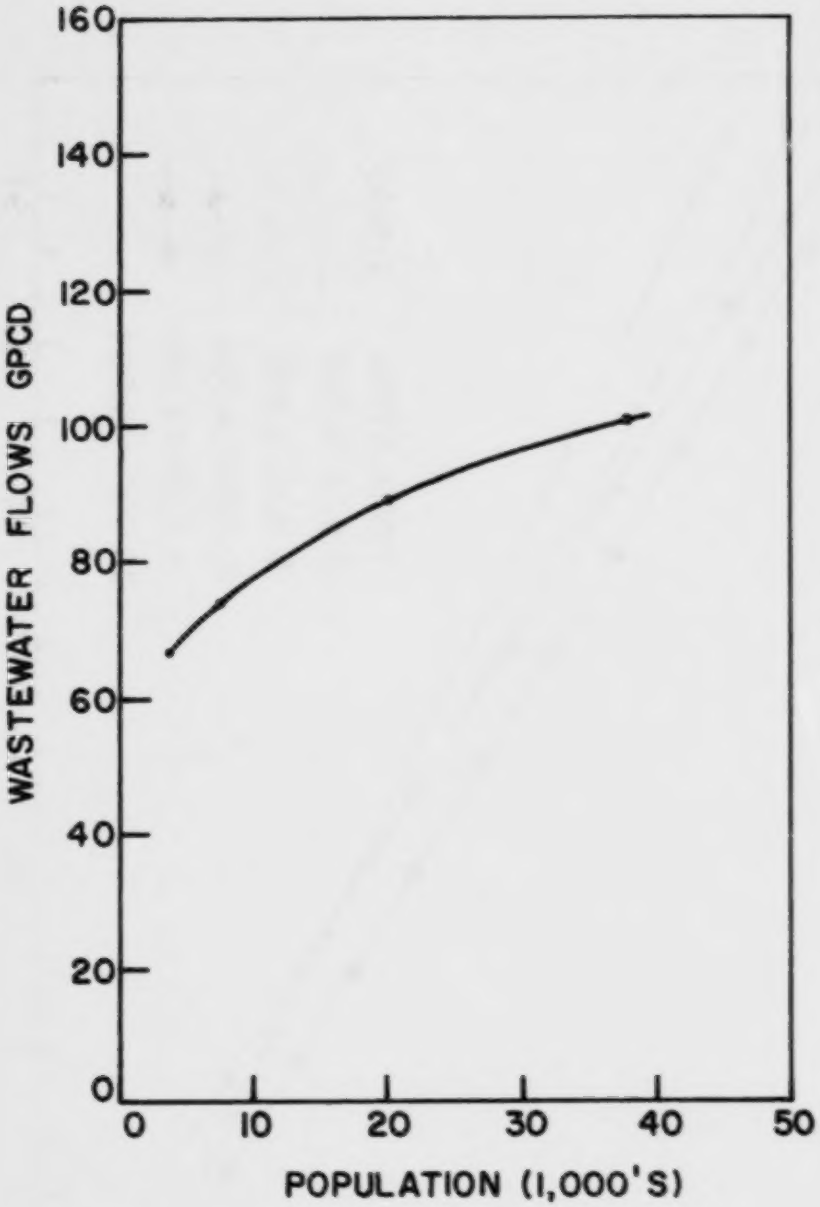

FIG. 1-1

EFFECTS OF POPULATION ON WASTEWATER FLOWS (Williamson 1970)

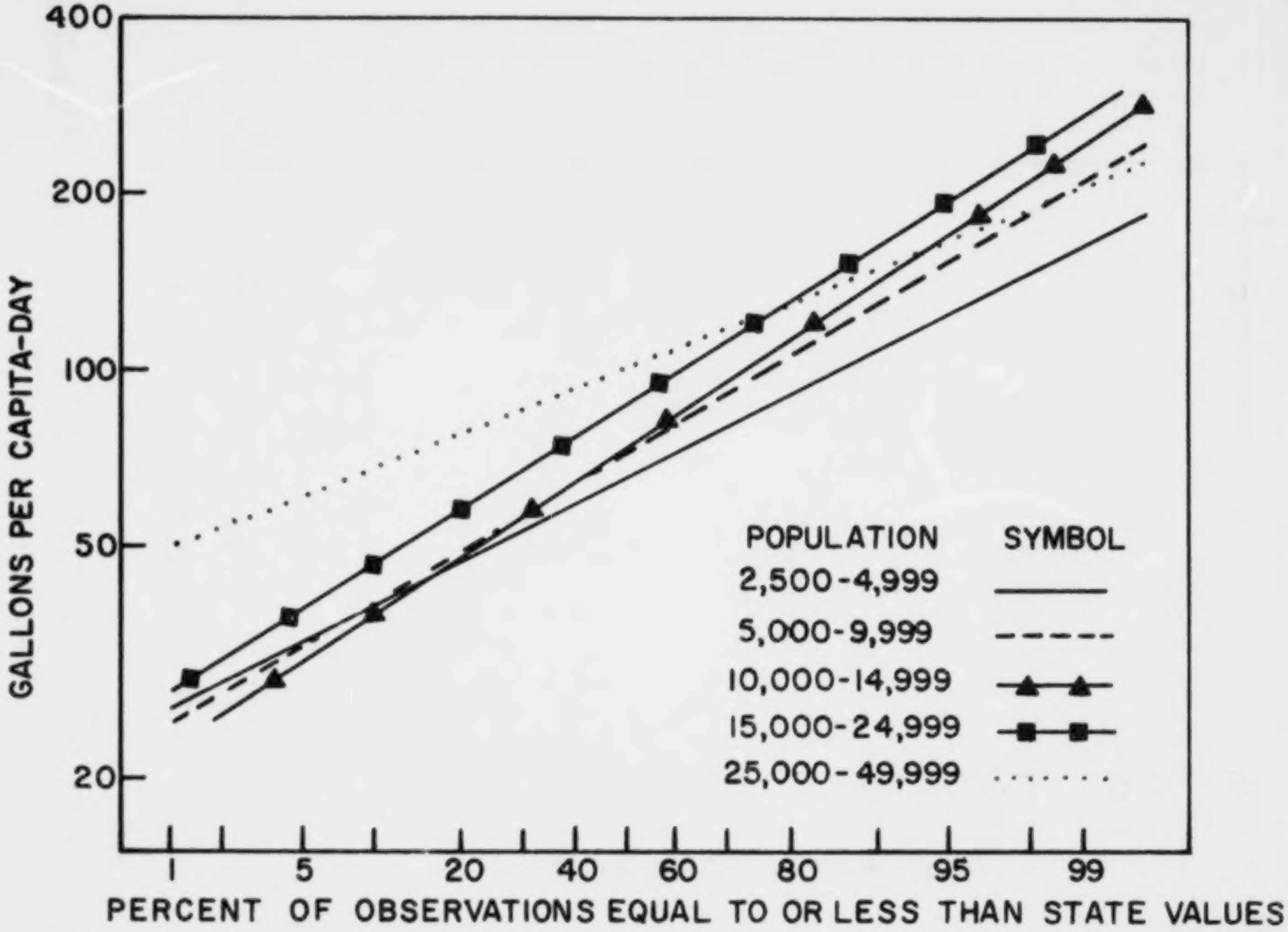

FIG.1-2
PROBABILITY ANALYSIS OF WASTEWATER FLOW
(Williamson 1970)

Water use expressed on a per capita basis varies over a wide range for different communities; therefore, the wastewater flow data can also be expected to have considerable variation. Some typical ranges of water use are illustrated in Table 1-2.

Table 1-2

Typical Water Use (Fair, 1971)

Use	Quantity, gallons/capita - day Normal Range	Average
Domestic	20-90	55
Commercial	10-130	20
Industrial	20-80	50
Public	5-20	10
Water Unaccounted for	5-30	15
Total	60-250	150

Water use in small communities which have very little industrial usage can be estimated at 85 - 100 gallons per capita per day (gpcd). The water use for larger communities in which commercial and industrial water usage is relatively high will reflect these water demands and the average water use will be increased to approximately 150 gpcd or greater depending on the type of industry.

The return wastewater flow into the collection system accounts for 60 to 70 percent of the water use. This percentage can be used to estimate the return flow when other information is not available. For example, the return flow for a small non-industrial community would be 60 to 70 gpcd if the water use was 100 gpcd. This flow range includes the average wastewater flow reported in Table 1-1 for a community with a population of 2500 to 4999 people. The range of return flow for a larger community with a water use of 150 gpcd would be 90 to 105 gpcd which includes the average return flow of 101 gpcd reported in Table 1-1 for communities with a population of 25,000 to 49,999. Therefore, water use records can be used to estimate the quantity and variation of wastewater flow necessary for the design of a wastewater collection system and treatment facility.

The quantity of wastewater also is markedly affected by rainfall. Runoff into combined systems is directly through the catch basins. However, rainfall may percolate into separate collection systems. Infiltration can significantly increase the flow and consequently the hydraulic load on the treatment facility.

Seasonal variations in flow and in rainfall are presented in Figure 1-3. The average monthly and maximum and minimum daily flows at the Govalle Treatment Plant in Austin, Texas during 1969 are included. There is some correlation between the monthly and maximum daily rainfalls and the average monthly and maximum daily flows. However, the intensity and duration of the rainfall have a dramatic affect on the flow to the treatment plant. A rainfall of low intensity and of long duration is much more likely to result in higher infiltration rates than a rainstorm of high intensity of a short duration. A statistical evaluation of the average daily flow data for 1969 is presented in Figure 1-4. These data indicate that municipal wastewater flow can be divided statistically into three separate groups, namely the average flow on Sundays, the average flow on work days, and the wastewater flow including infiltration. The average flow at the Govalle Plant on Sunday is 20 MGD and on work days 23.5 MGD. The data in Figure 1-4 indicate that an average dry weather flow of 25 MGD will occur 90 percent of the time. However, a flow of 30 MGD will occur 90 percent of the time during periods of rainfall.

A daily variation in flow during dry weather flow (July 16-17, 1970) and during a rainy period (May 20-21, 1970) are presented in Figure 1-5. The difference in the average daily flow is 14 MGD and is attributed to infiltration. If this average rate of infiltration is subtracted from the flow recorded for the May 20-21, 1970, curve, the resulting curve will be very similar to the dry weather flow curve. The ratio of peak daily flow to average daily flow ranges from 1.27 to 1.45, respectively, for the wet weather flow and for the dry weather flow curves. The average flow during the dry weather for the time period 10:00 a.m. to 10:00 p.m. is 28 MGD.

The hydraulic design of a municipal wastewater treatment plant is markedly affected by the peak flow as well as the variation in flow during the course of the day. Therefore, the flow variations must be included in any evaluation of treatment plant performance.

WASTEWATER COMPOSITION

The composition and concentration of the constituents of wastewater are affected by the flow, infiltration of rainfall, the type of industrial development and the extent to which the sanitary collection system is separated from the stormwater system. Some typical data relating flow to the concentrations of BOD and suspended solids (SS) are presented for various cities in Figures 1-6 through 1-11. The data illustrated in Figure 1-6 were

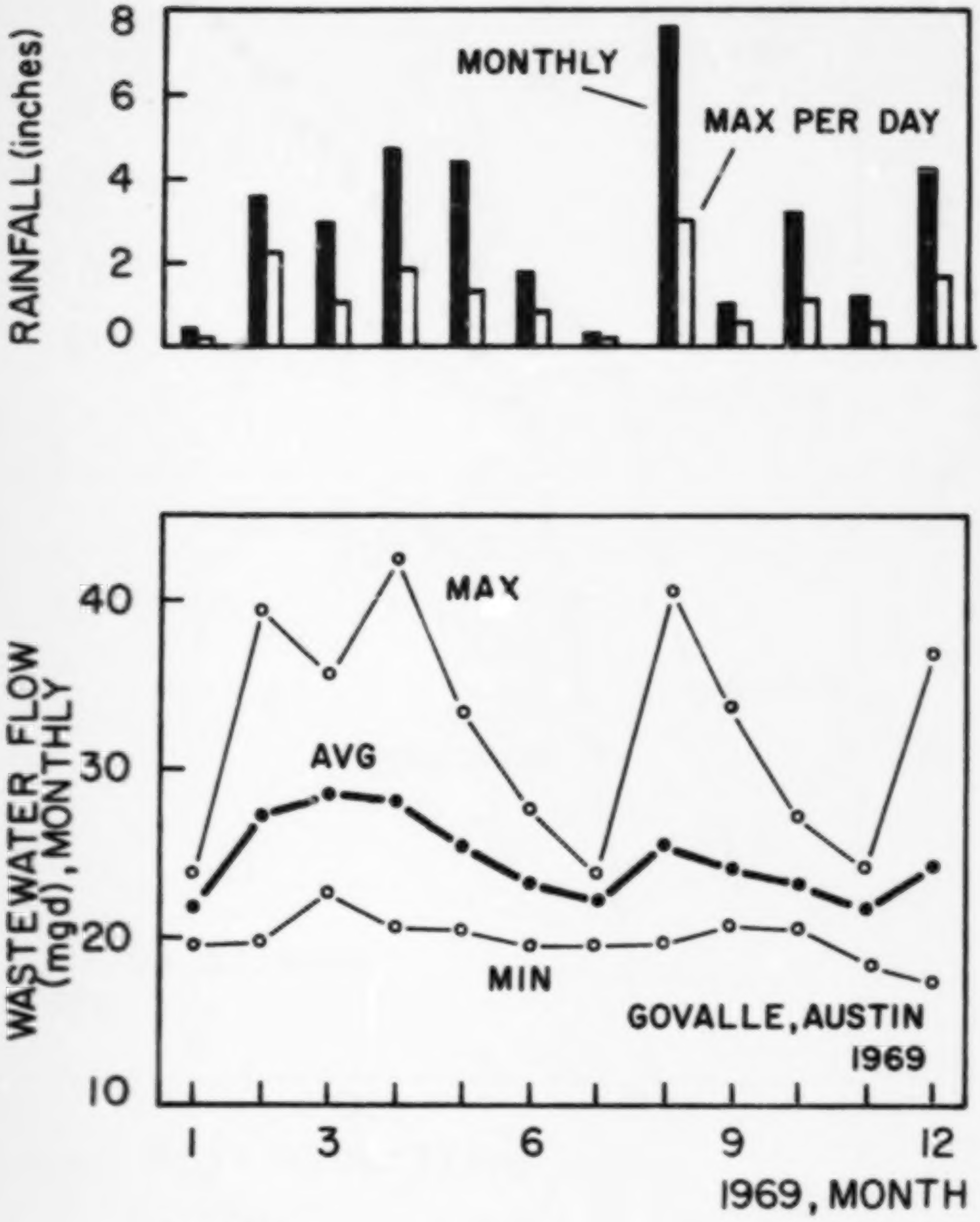

FIG. 1-3
RAINFALL AND WASTEWATER FLOW

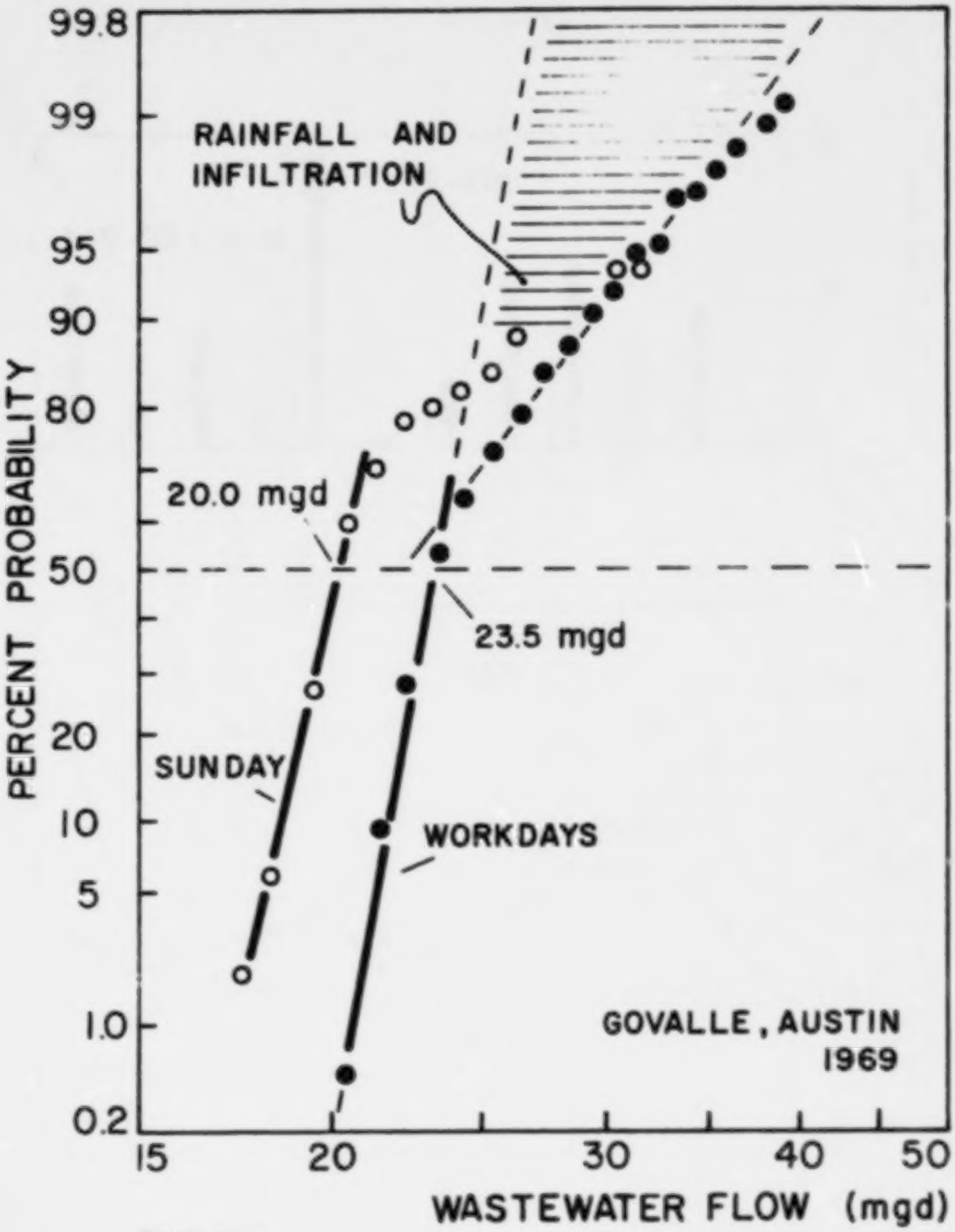

FIG. 1-4

PROBABILITY OF WASTEWATER FLOW

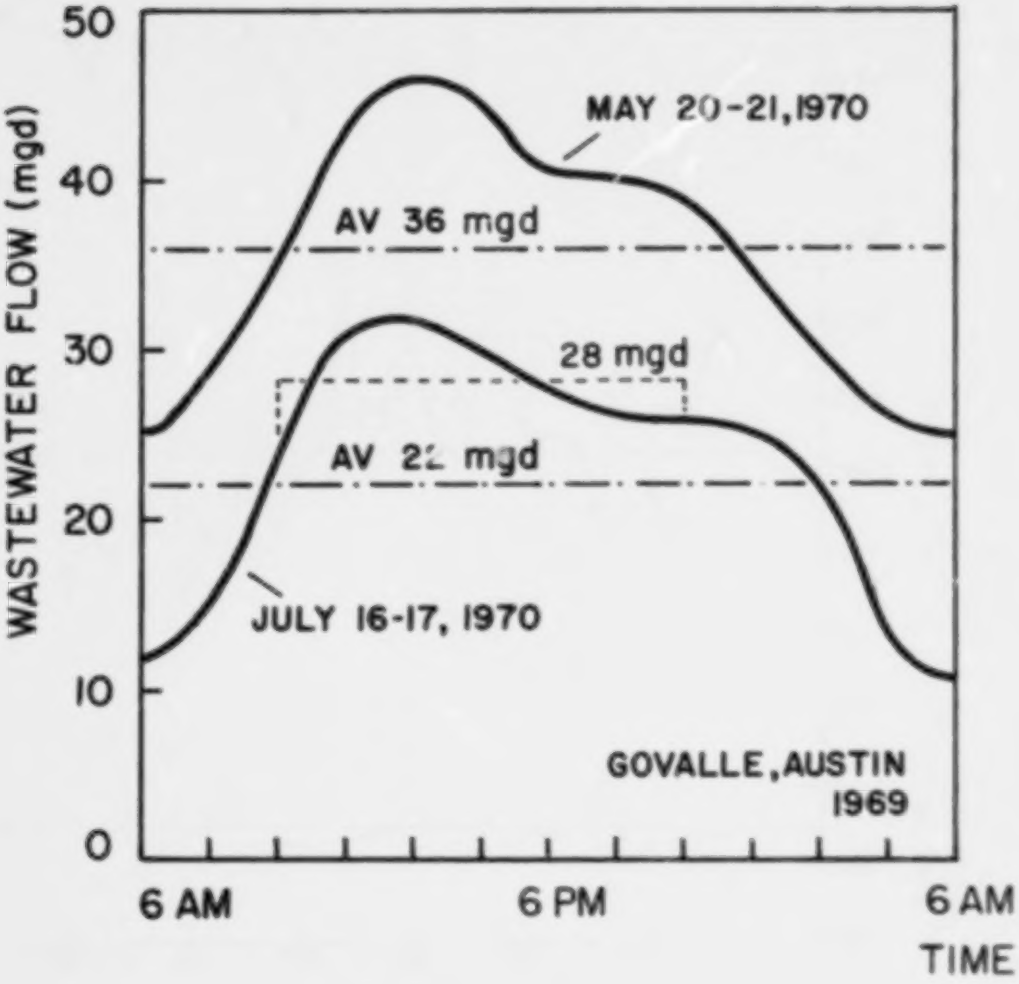

FIG.1-5
WASTEWATER FLOW DISTRIBUTION
GOVALLE PLANT, AUSTIN

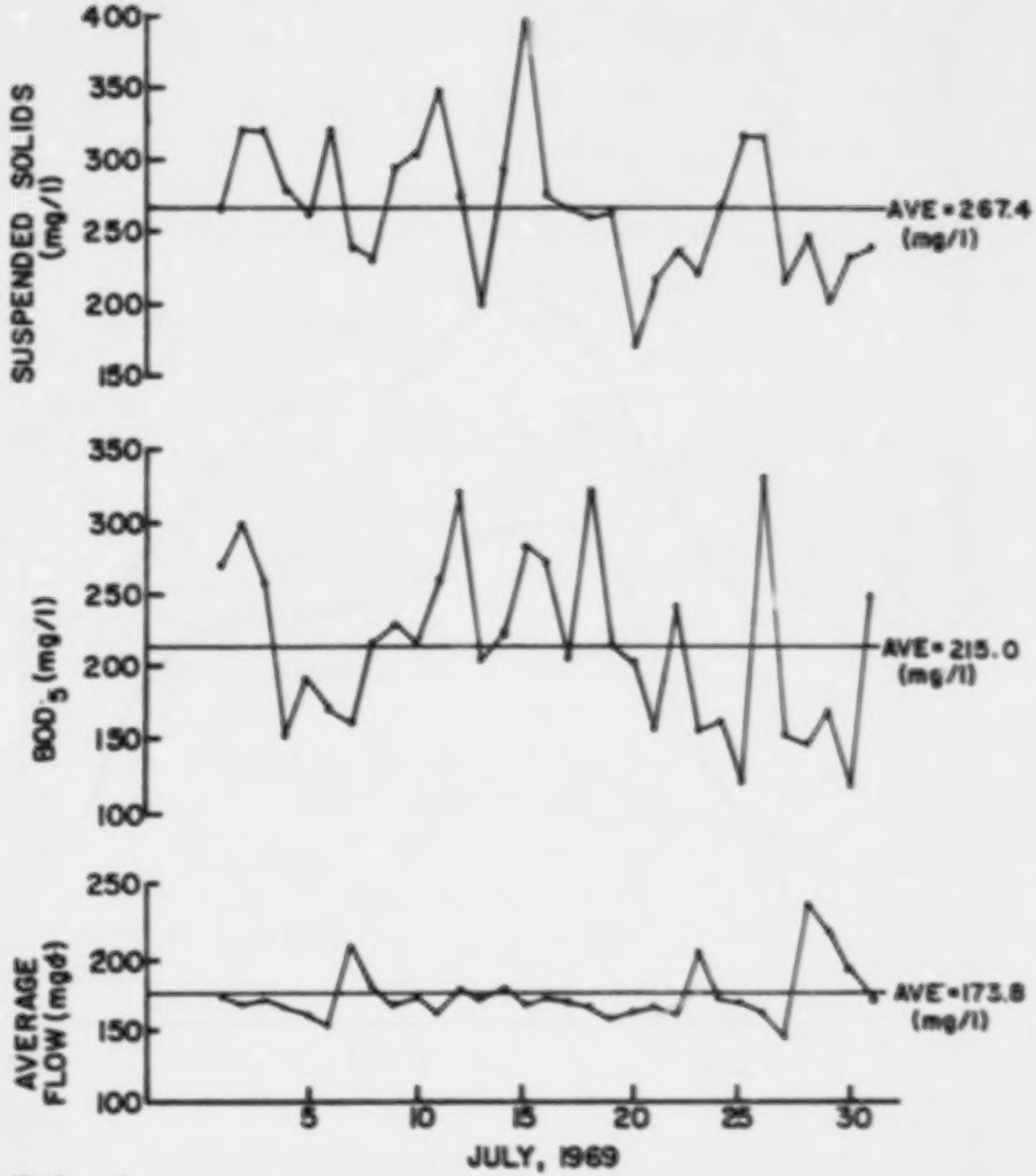

FIG. 1-6
DAILY VARIATIONS IN WASTEWATER FLOW & COMPOSITION PHILADELPHIA NORTHEAST WATER POLLUTION CONTROL PLANT – JULY, 1969

reported for July, 1969, by the City of Philadelphia which has a combined collection system which accepts industrial wastes. The daily variations in the concentrations of BOD and SS are more pronounced than the daily variations in the wastewater flow. It is difficult to attempt to correlate the relationship of workday or Sunday to wastewater flow and composition since the frequency and quantity of rainfall is not reported.

The monthly variations in the wastewater flow and composition for the Northside, Calumet, and Southwest Plants of the Chicago Metropolitan Sanitary District are presented in Figures 1-7, 1-8, and 1-9, respectively. Domestic and industrial wastewaters are collected in a combined system; therefore, the data represent the influence of stormwater runoff. These data indicate that the average concentrations of BOD and SS in the untreated wastewater entering the Southwest Plant which receives most of the industrial wastes is about ten times higher than that entering the other two plants.

The effects of industrial wastes and stormwater on the composition of the wastewater are indicated by the wide range of concentrations included between the monthly maximum and minimum concentrations reported. The monthly average flow, BOD and SS are relatively constant for the Northside and Calumet Plants compared to those for the Southwest Plant.

The average monthly data for the two treatment plants in Fort Worth, Texas, are presented in Figure 1-10. The collection systems are separate. The flow rate, BOD and SS are relatively constant throughout the year. The variations in the concentrations of SS and BOD are attributable to industrial wastes discharged into the collection system.

The data for the City of San Antonio, Texas, are presented in Figure 1-11. The variations in wastewater flow can be related to the infiltration of rainfall as indicated by the data. The decrease in BOD as the flow rate increased indicated the diluting effect of infiltration. The suspended solids concentration remained relatively constant indicating that the infiltration did not dilute the solids. However, the increased flow resulting from infiltration may have resuspended some of the heavier solids which may have settled in the collection system and offset the diluting effect.

BOD Loading

The interrelationship of flow, concentration of BOD in the wastewater and the total organic load to the treatment plant can be used to evaluate the dilution which generally results from infiltration of rainwater. The data presented in Figure 1-12 indicate the variations in wastewater flow, concentration of BOD and BOD load expressed in thousand pounds/day for the Govalle

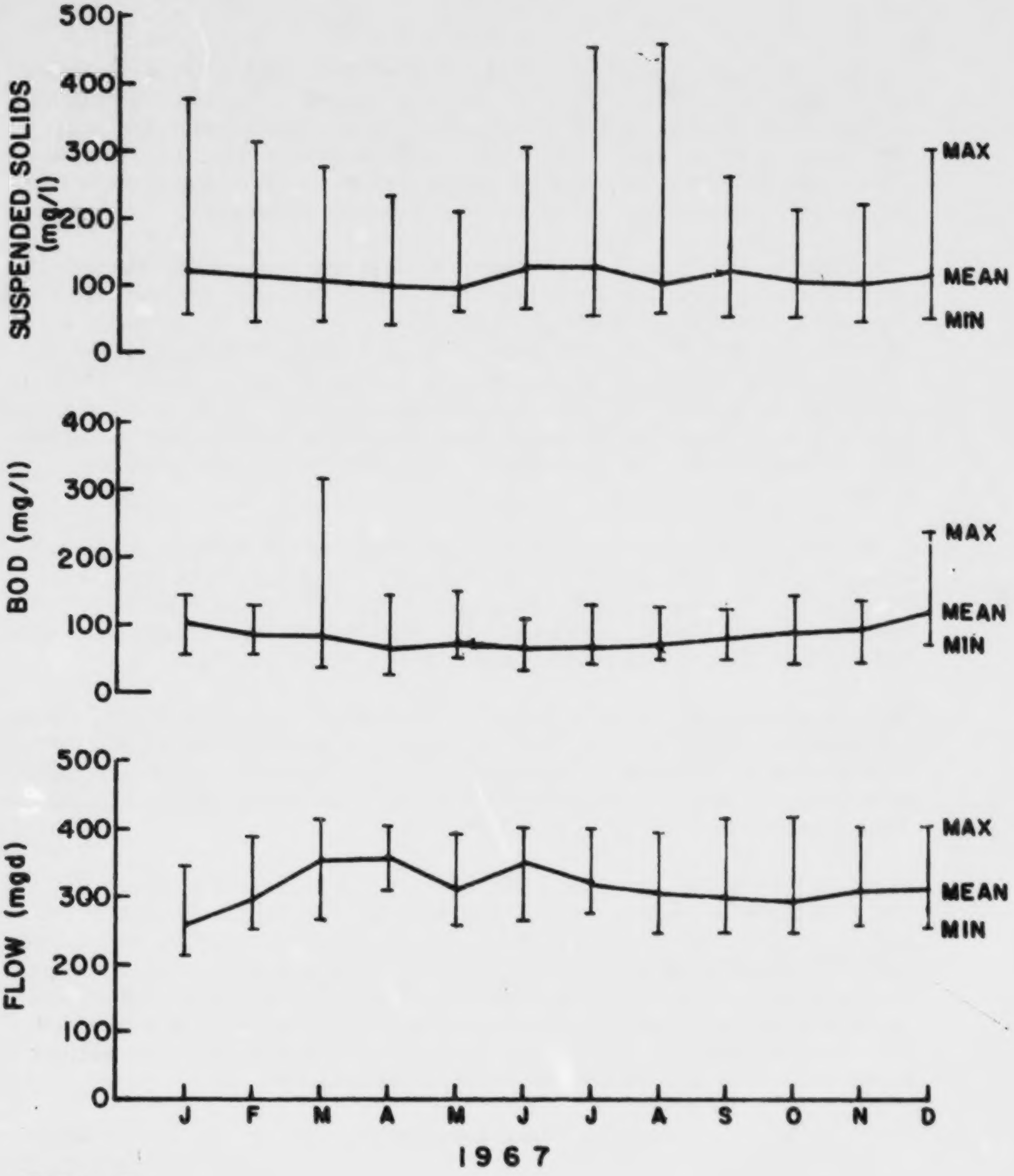

FIG.I-7
MONTHLY VARIATIONS IN WASTEWATER FLOW & COMPOSITION (Northside plant)
CHICAGO METROPOLITAN SANITARY DISTRICT

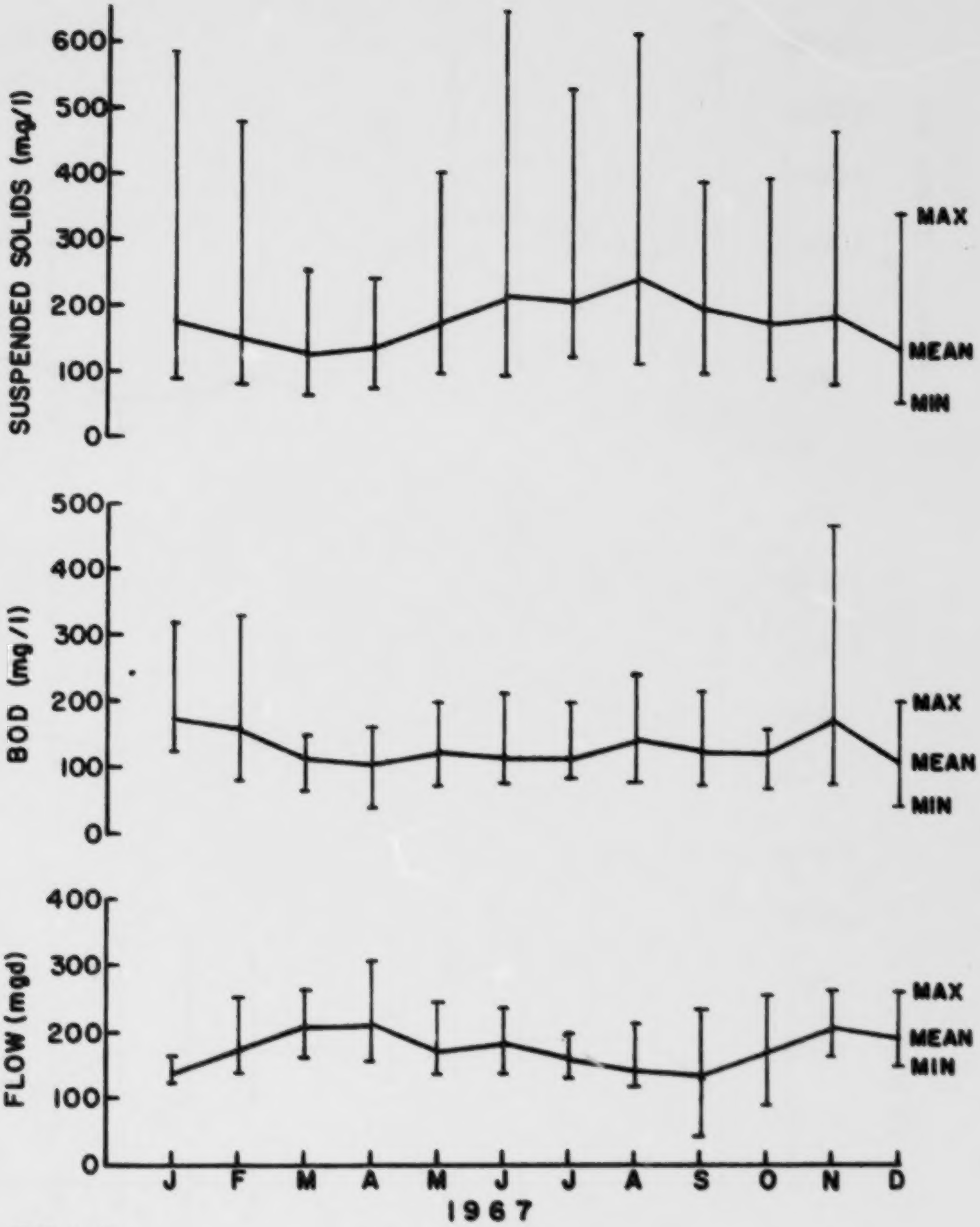

FIG.1-8
MONTHLY VARIATIONS IN WASTEWATER FLOW COMPOSITION
CHICAGO CALUMET PLANT - 1967

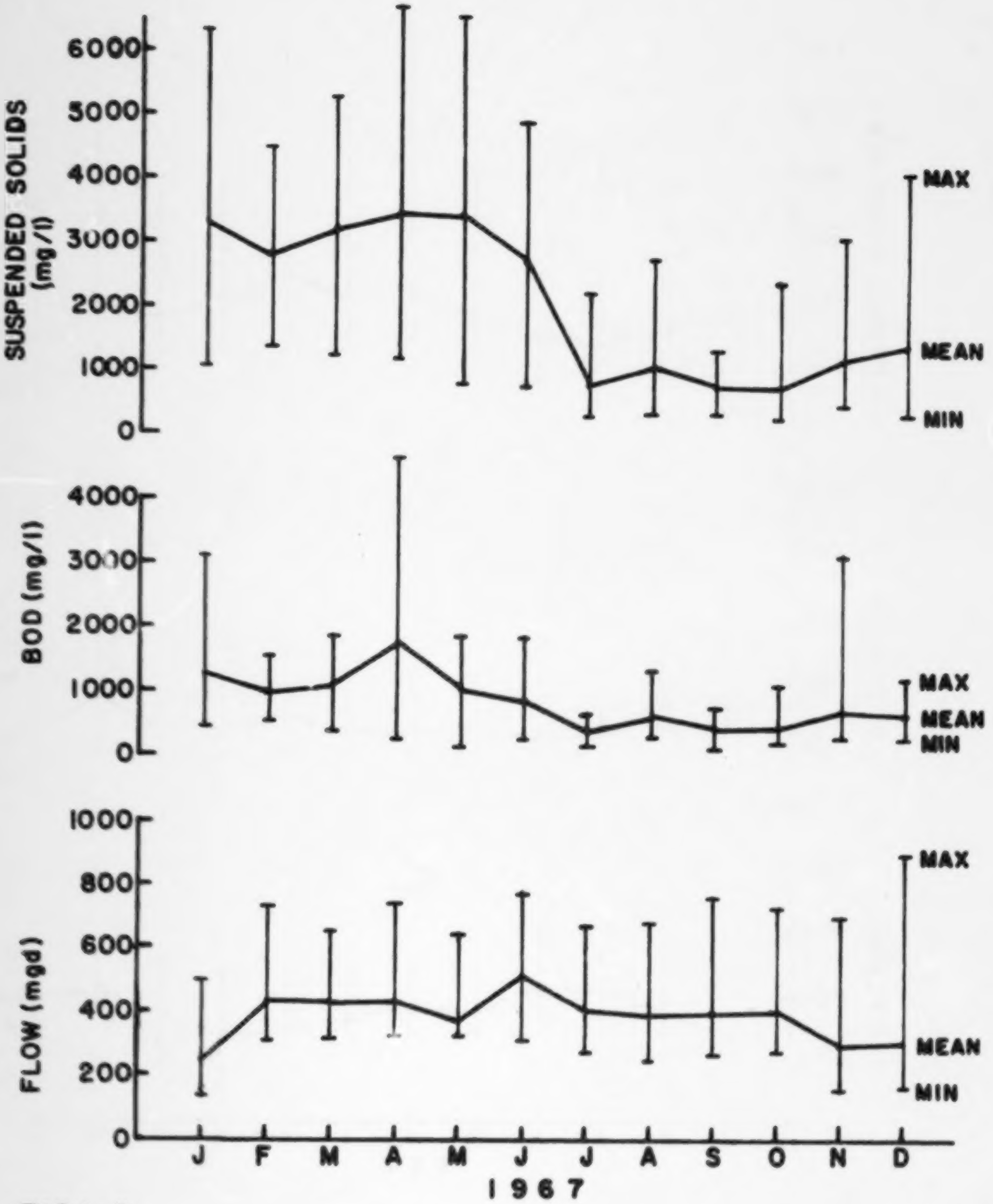

FIG.1-9
MONTHLY VARIATIONS IN WASTEWATER FLOW & COMPOSITION CHICAGO SOUTHWEST PLANT-1967

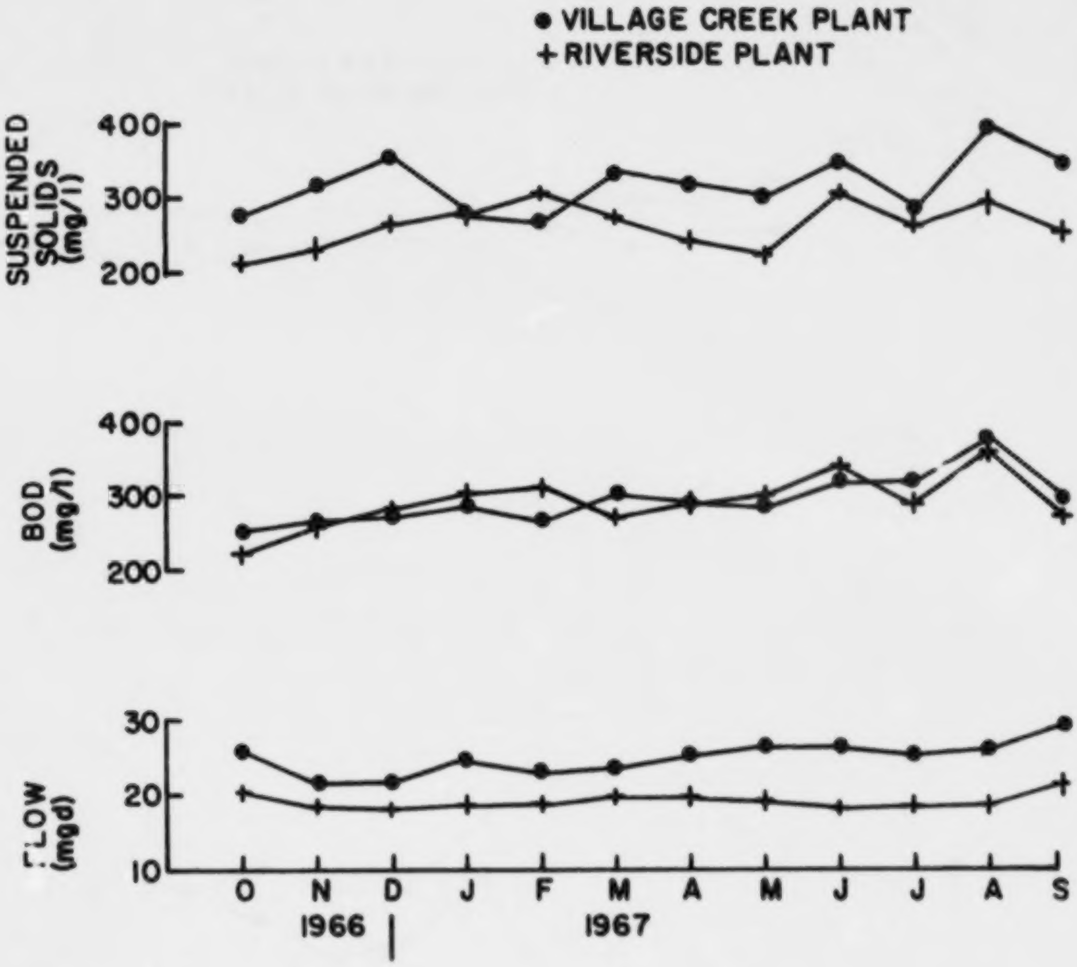

FIG.1-10
MONTHLY VARIATIONS IN WASTEWATER FLOW &
COMPOSITION - FORT WORTH, TEXAS

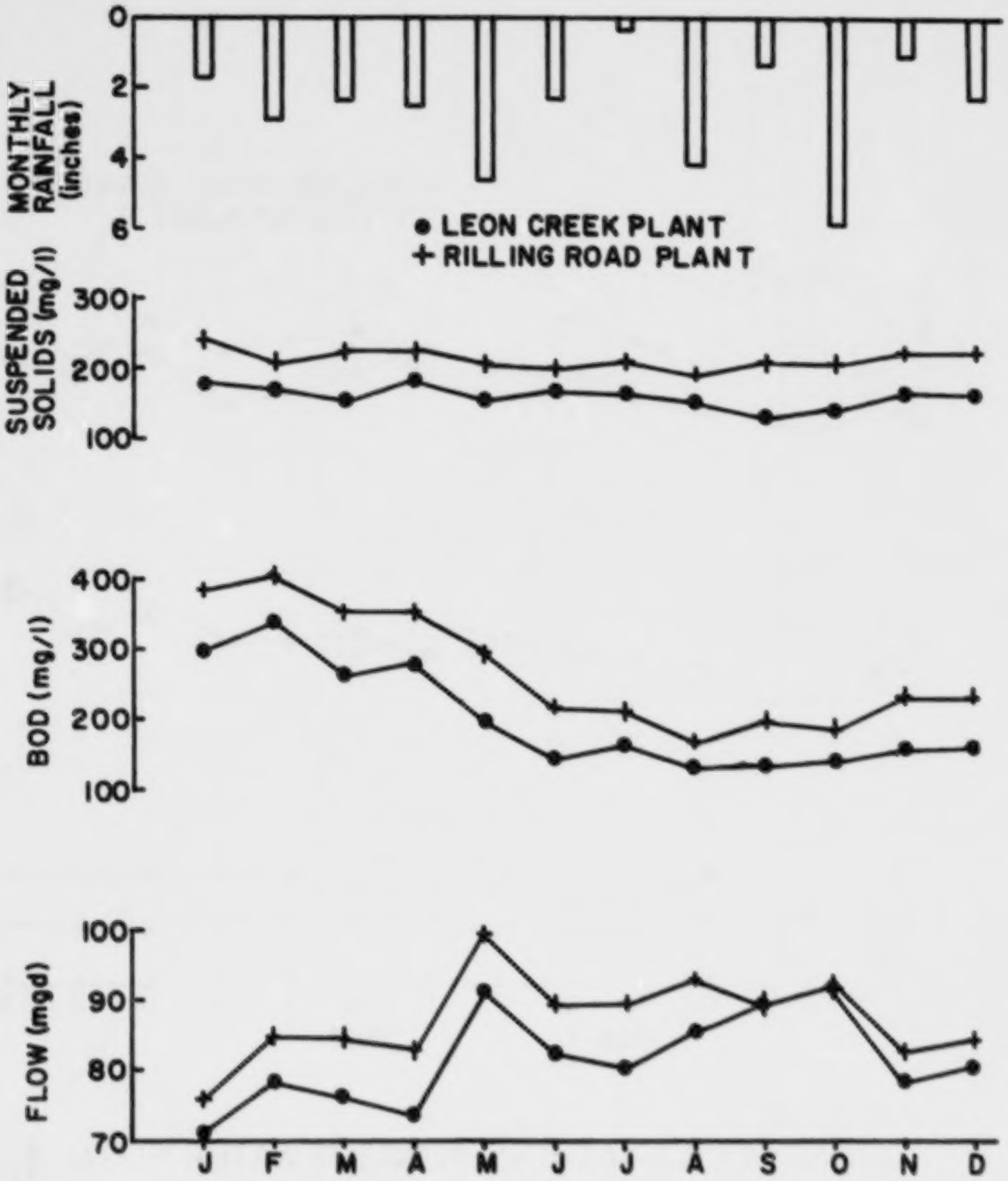

FIG.1-11
MONTHLY VARIATIONS IN WASTEWATER FLOW &
COMPOSITION – SAN ANTONIO, TEXAS (1966)

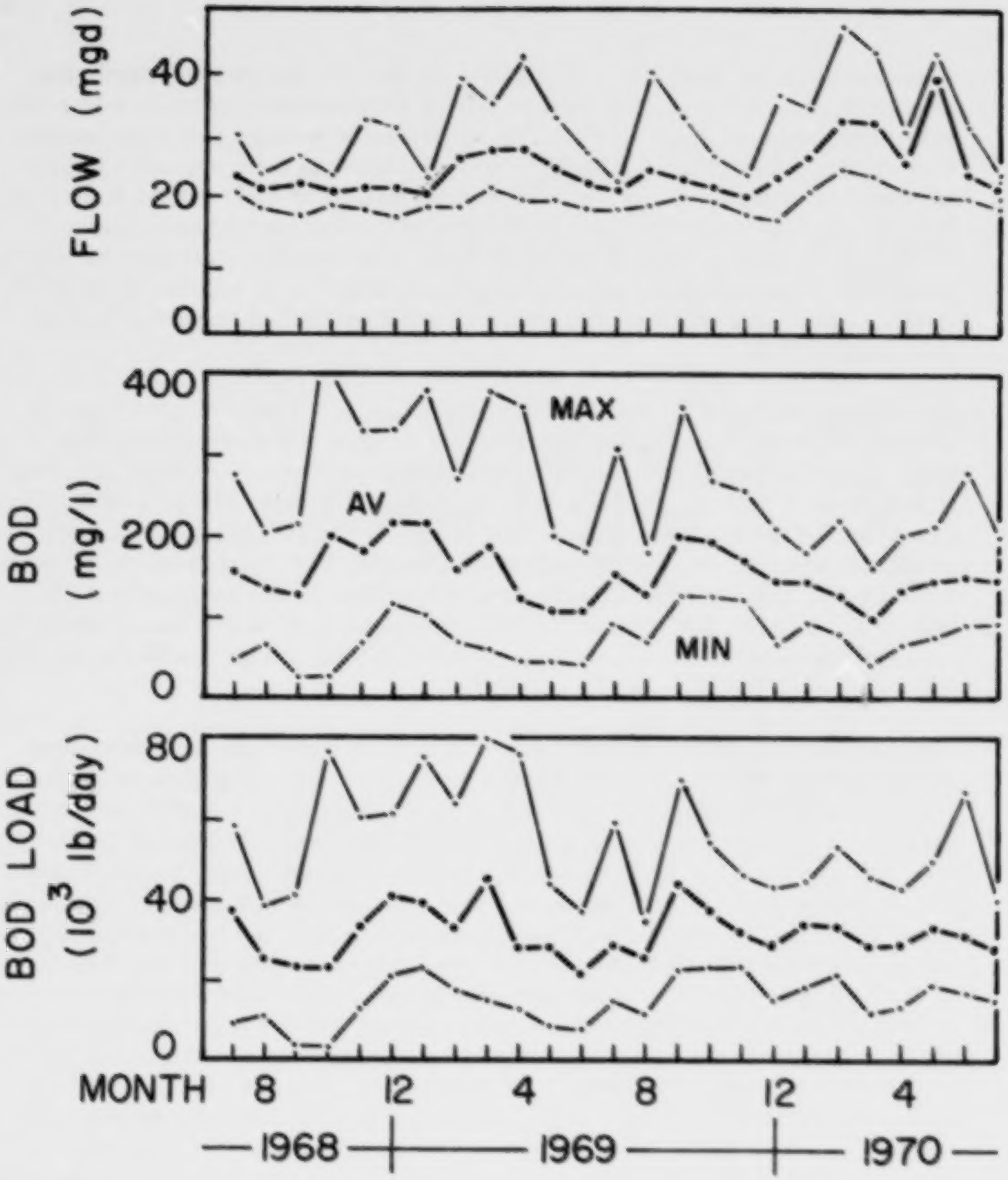

FIG.1-12

LOADING CHARACTERISTICS

MONTHLY AVG, MAX, MIN GOVALLE

Wastewater Treatment Plant. These data are the monthly average, the maximum, and the minimum values which occurred during the time period, July, 1968, through June, 1970. The variation in wastewater flow ranged from slightly less than 20 MGD to over 45 MGD and represents about a two-fold variation. However, the concentration of BOD varied from less than 50 mg/l to over 400 mg/l while the BOD load ranged from about 10,000 to 80,000 pounds of BOD per day. The ratio of maximum to minimum BOD concentration and load are about 8 to 1. Therefore, total BOD load is more dependent on the concentration of BOD than on the rate of wastewater flow.

Statistical analysis of the BOD concentration and load are presented in Figures 1-13 and 1-14, respectively. These data are described by a logarithmic or geometric distribution. There is some deviation from the linearity at a BOD concentration of less than 100 mg/l and at a BOD loading of less than 15,000 lb/day. The slopes of the two lines are almost identical and this similarity indicates that the BOD load is affected much more by the variation in BOD concentration than in the wastewater flow. The data indicate that during the two year period of record the average BOD into the Govalle Wastewater Treatment Plant was 155 mg/l and the average BOD load was 30,000 pounds per day.

The relationship of wastewater flow, BOD concentration, and BOD load during a 24-hour period is presented in Figure 1-15. The data in Figure 1-15 are based on the analyses of four-hour composite samples which were prepared by mixing hourly samples. These data were collected during July 16, 1970, and July 17, 1970, which represents a period of dry weather flow. The concentration of BOD is highest for the sample collected during 12:00 noon to 4:30 p.m. period and decreases for each of the following four-hour composite samples. The sample collected between 8:30 p.m. and 12:30 a.m. is the lowest. The flow rate also is highest at noon and decreases steadily until 4:30 a.m. The BOD load is maximum during the four-hour period of 12:30 and 4:30 p.m., decreases steadily and reaches a minimum between 4:30 and 8:30 a.m. The peak BOD load is 2500 pounds of BOD per hour while the average load was 1400 pounds per hour. The ratio of the peak to the average was 1.7. The average load during the day-time period from 12:30 p.m. to 8:30 p.m. was approximately 2,200 pounds of BOD per hour. The ratio of this average load to the 24-hour average load is about 1.5. During this eight-hour period, about half of the total load arrives at the plant.

The data indicate that transient loading conditions should be considered in the design of biological treatment facilities. The BOD load to the plant expressed in pounds per hour provides a better estimate of the load on the facility than the variations in flow and BOD concentration.

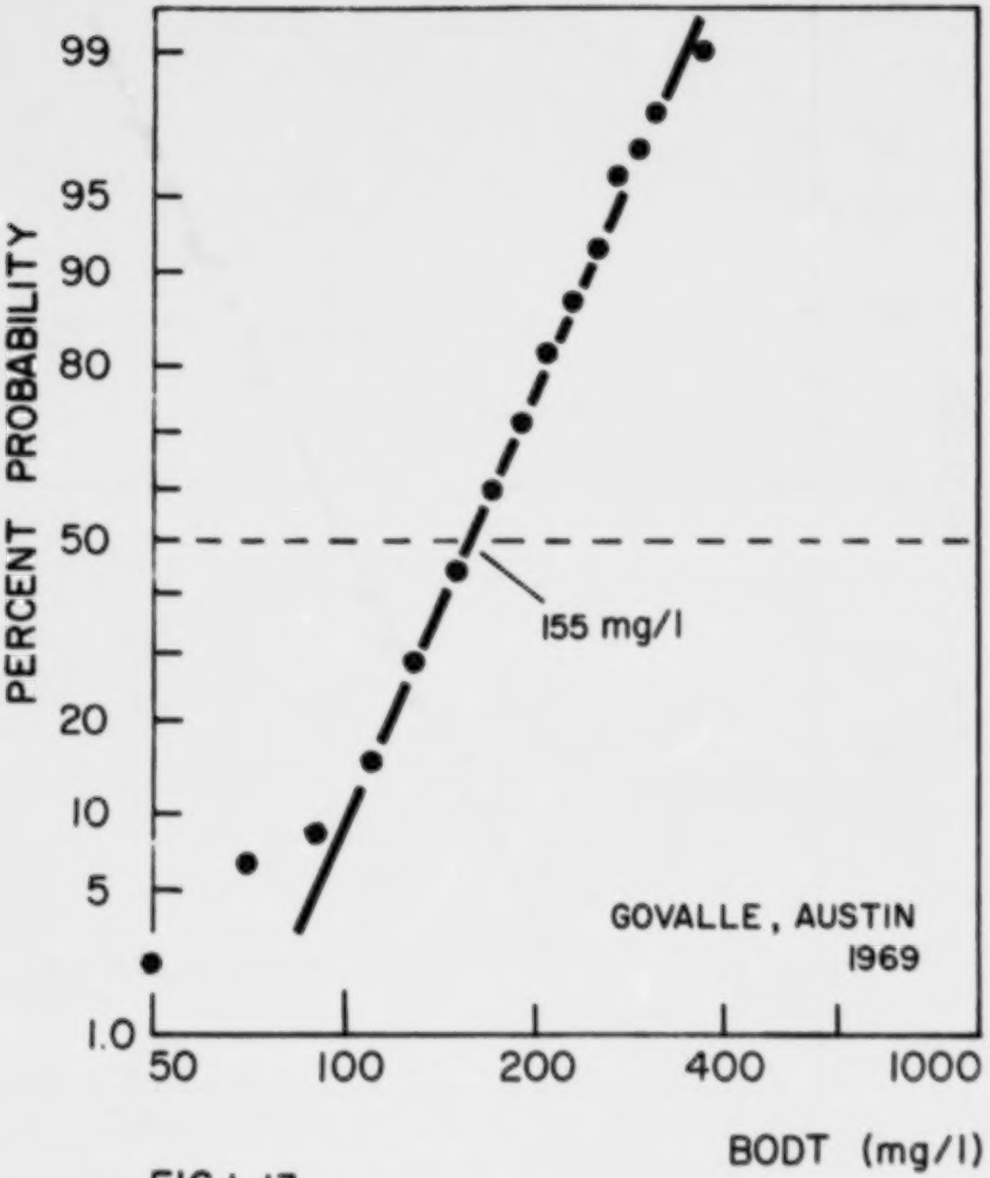

FIG.1-13

PROBABILITY OF BOD INFLUENT

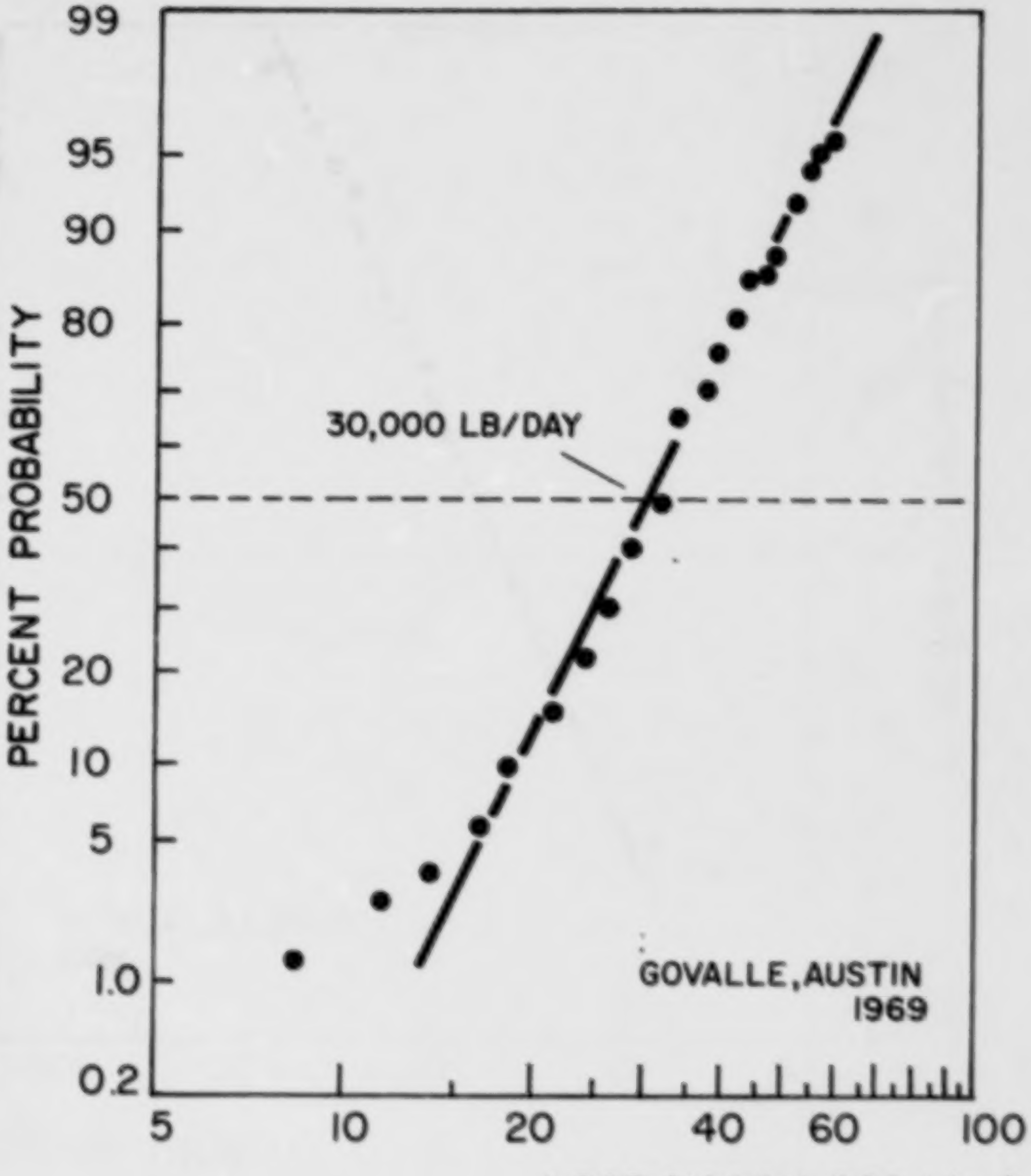

FIG. 1-14

PROBABILITY OF BOD LOAD

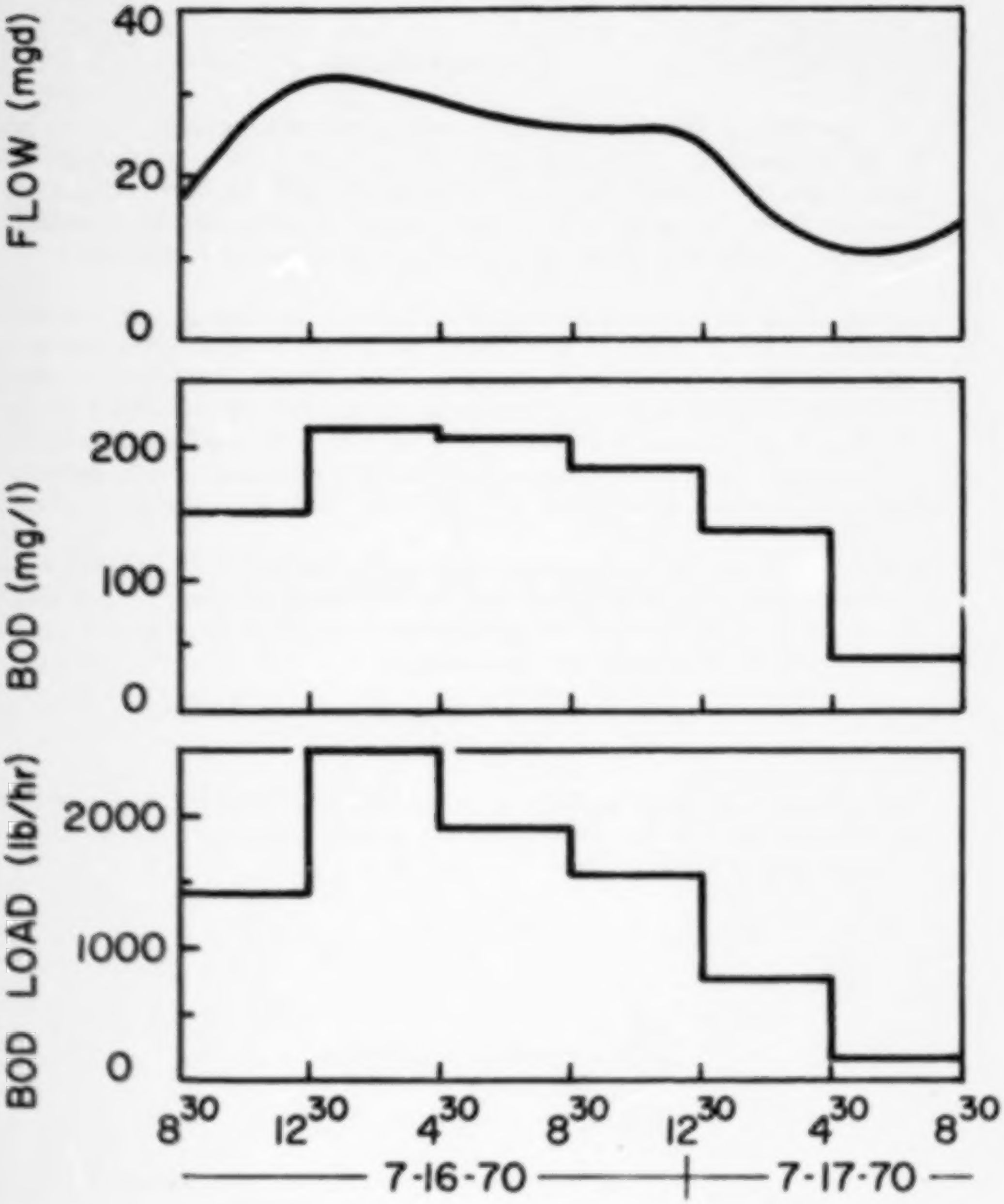

FIG.1-15

BOD AND LOAD DISTRIBUTION

GOVALLE, AUSTIN, TEX

Suspended Solids

The quantity of sludge requiring handling and disposal is directly related to the suspended solids concentration in wastewater. Statistical evaluation of the suspended solids data for the Govalle Treatment Plant during 1969 is shown in Figure 1-16. These data are described by a normal distribution. The mean suspended solids concentration is 155 mg/l.

A portion of the volatile suspended solids are biodegradable. Inefficient removal of the volatile solids adds a BOD load to and increases the oxygen requirements of biological treatment. The relationship between the volatile suspended solids and total suspended solids for the Govalle and Hyperion treatment plants are presented in Figure 1-17. This linear relationship between the volatile fraction and the total suspended solids indicates that the volatile fraction accounts for 86.5 per cent of the suspended solids.

A fraction of the volatile suspended solids are not biodegradable, and therefore accumulate in the biological treatment system. Smith and Eilers (1969) suggested that the biodegradable fraction of the volatile suspended solids can be calculated by Equation 1-1:

$$X_{ovb} = X_{ov} \left(\frac{\text{ultimate BOD of Suspended Solids}}{\text{COD of Suspended Solids}} \right) \qquad 1\text{-}1$$

The ratio of the five-day BOD to the ultimate BOD of the suspended solids is assumed to be 0.9. Therefore, the nonbiodegradable suspended solids can be calculated from Equation 1-2:

$$X_{on} = X_{ov} \left(1 - \frac{\text{BOD}}{0.9\ \text{COD}}\right) + (X_o - X_{ov}) \qquad 1\text{-}2$$

in which:

X_o = total suspended solids (mg/l)

X_{ov} = volatile suspended solids (mg/l)

X_{ovn} = non biodegradable volatile suspended solids (mg/l)

X_{on} = non biodegradable solids (mg/l)

BOD = five-day BOD of the suspended solids (mg/l)

COD = the COD of suspended solids (mg/l)

The relationship between the volatile suspended solids and nonbiodegradable volatile suspended solids in the untreated wastewaters is shown in Figure 1-18.

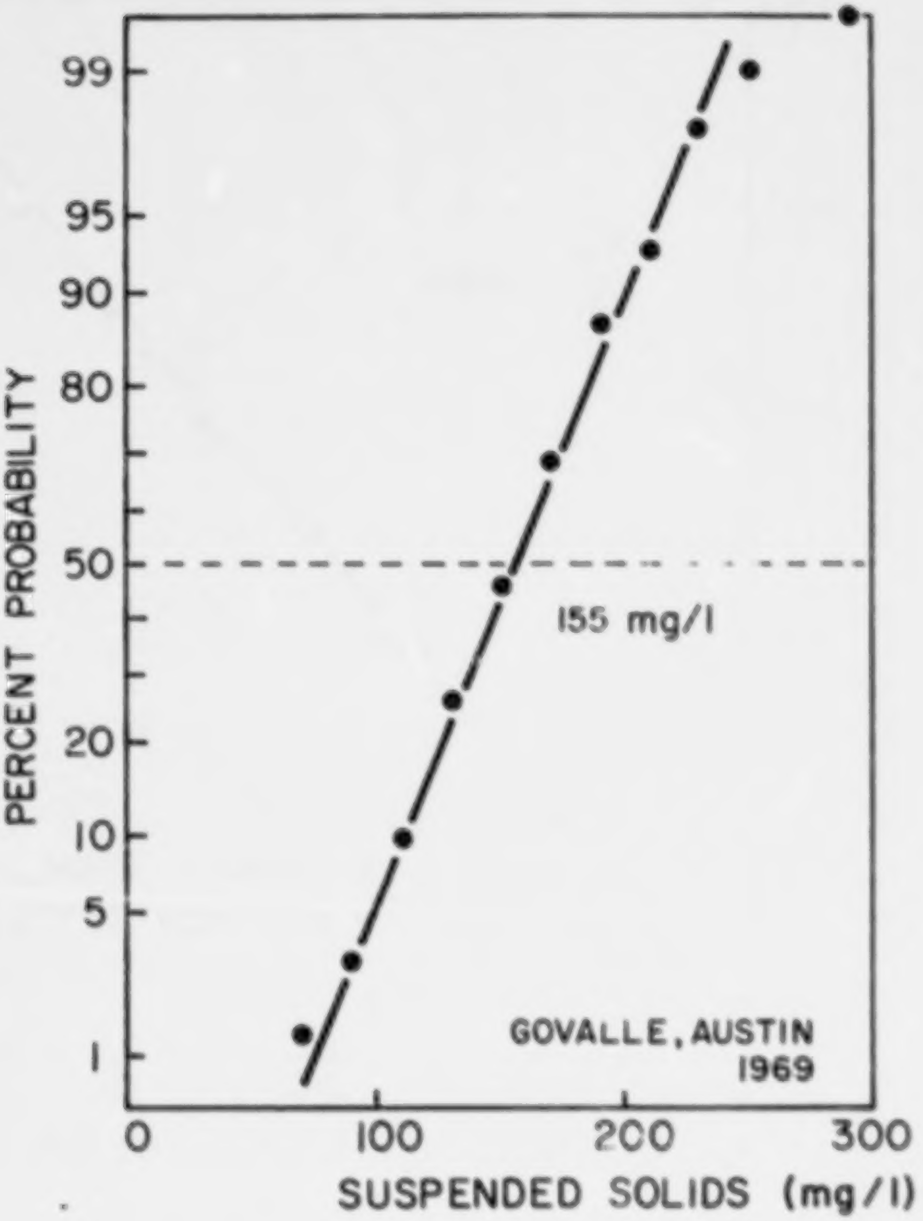

FIG.1-16
PROBABILITY OF SUSPENDED SOLIDS

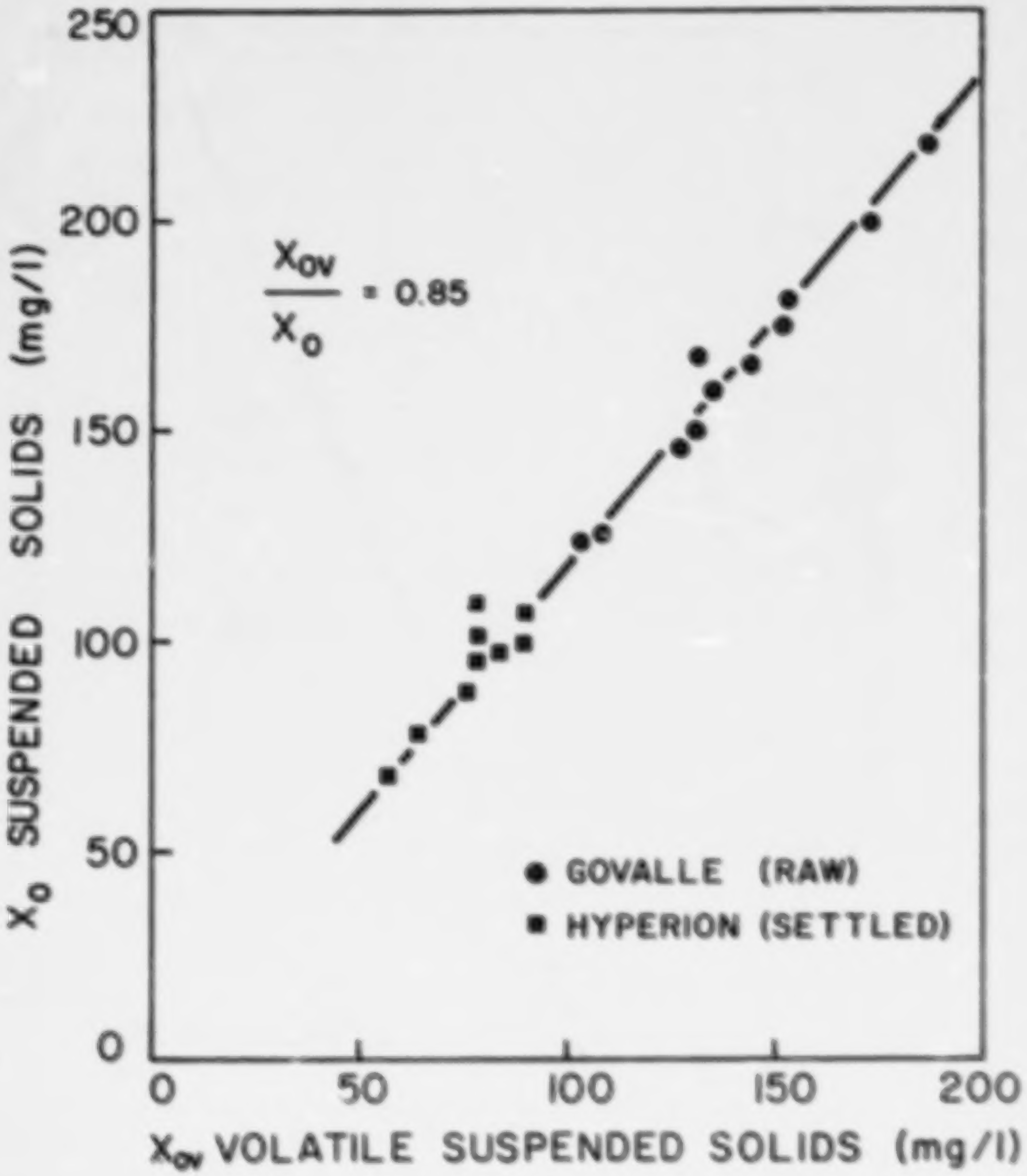

FIG. 1-17
TOTAL AND VOLATILE SUSPENDED SOLIDS OF RAW SEWAGE

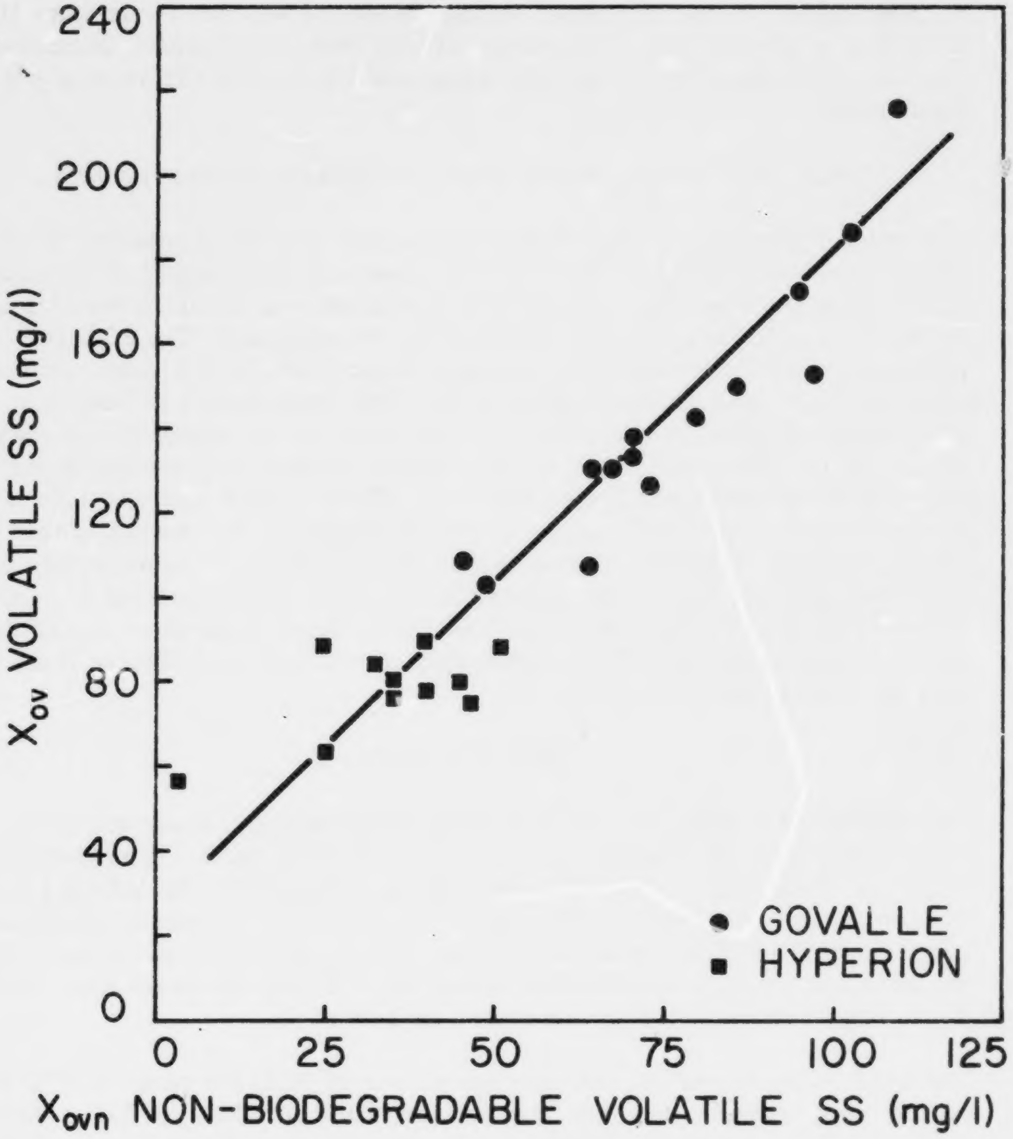

FIG. 1-18

NON-BIODEGRADABLE VOLATILE SUSPENDED SOLIDS IN RAW WASTEWATER

The data indicate that the concentration of nonbiodegradable volatile solids increases as the volatile suspended solids concentration increases. The relationship between the nonbiodegradable suspended solids and the total suspended solids concentration in the untreated wastewater is more linear as shown in Figure 1-19. The slope of this line is 0.6 which indicates that about 60 per cent of the total suspended solids in the sample are nonbiodegradable.

Parameters for Estimating Organic Contents of Wastewaters

The organic fraction of municipal wastewaters can be characterized by organic carbon in conjunction with the classical BOD and COD information. These three parameters are used interchangeably at times; however, each measures a different characteristic of the wastewater. The BOD indicates the fraction of the wastewater which is degradable by the seed microorganisms during a specific incubation time. The more slowly or less readily degradable material is generally not reflected in the standard five-day BOD test. On the other hand, the COD includes organic and inorganic materials which exert a chemical oxygen demand. Much of this material may be nonbiodegradable. The TOC represents an estimate of the biodegradable and nonbiodegradable carbon in the wastewater material. A relationship or ratio between any two of the parameters must be developed for the particular wastewater in order to effectively use these three parameters in the design of the treatment plant. The relationship of the BOD and TOC is linear and may be expressed by Equation 1-3:

$$\mathrm{BOD} = a\,(\mathrm{TOC}) - b \qquad 1\text{-}3$$

The constants a and b can be evaluated graphically. However, the number of uncertainties associated with the BOD analysis make a true evaluation difficult. The value of these constants is different for the influent and the effluent samples since the effluent of a wastewater treatment plant contains the less readily biodegradable material. The coefficient of variance for the data observed at the Govalle Wastewater Treatment Plant and those presented by Wuhrmann (1964) are shown in Table 1-3.

The data indicate that at low concentrations of BOD the ratio of BOD to TOC varies much more widely than the similar ratios for COD to TOC. Some of this variation may be attributable to the limitations of the BOD analysis at low substrate concentrations. Wuhrmann (1964) indicated that the BOD analysis represents oxygen uptake in a limited period of time by an unknown number of organisms and an unknown substrate. In spite of the shortcomings of the BOD determination, the design information presented in these guidelines are based on the BOD since sufficient TOC and COD data and experience on which to base the design of a municipal treatment plant are not available.

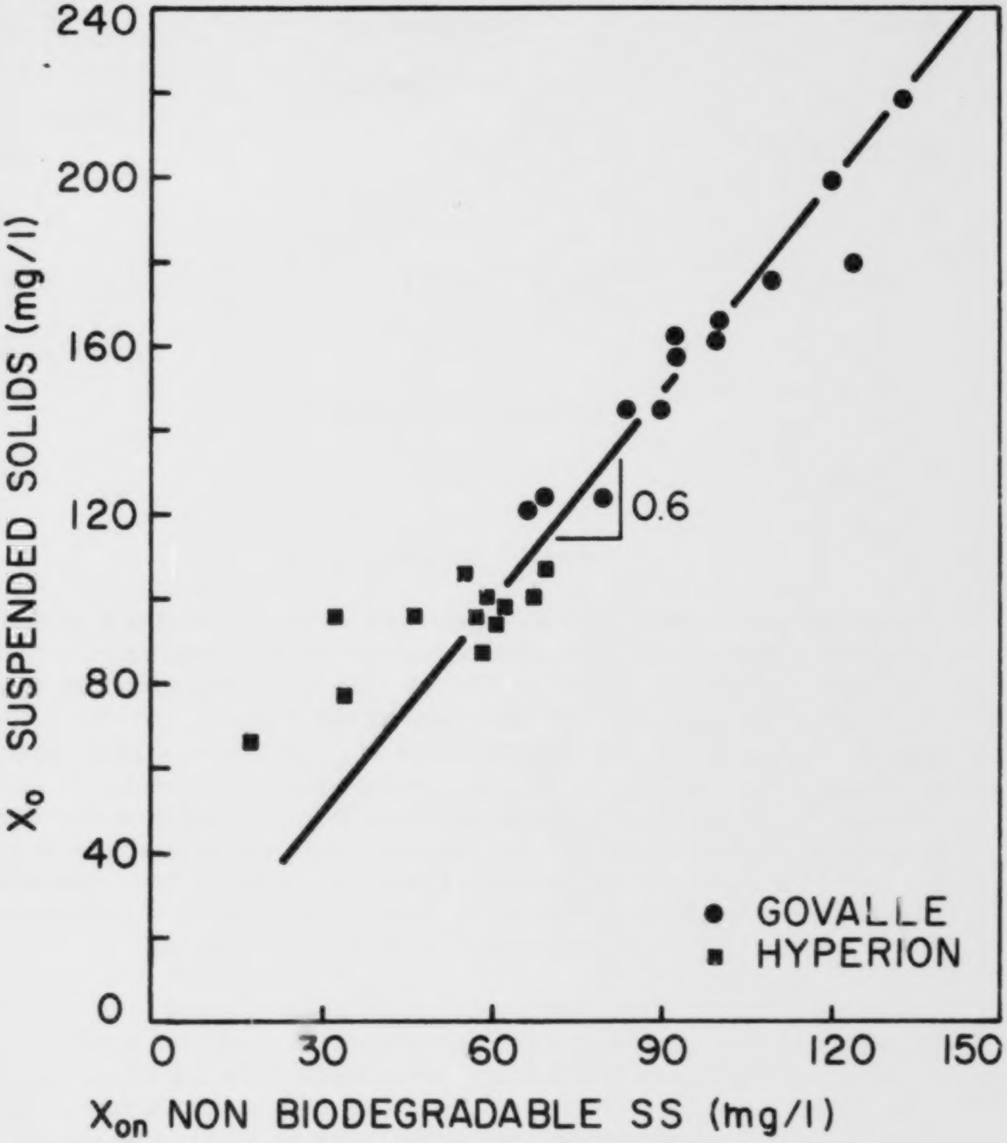

FIG.1-19
NON BIODEGRADABLE SUSPENDED SOLIDS
IN RAW WASTEWATER

Table 1-3

Ratios of the Coefficients of Variance

Ratio	Influent	Effluent
	Govalle Experiments	
BODT/TOCT	0.91	1.38
BODF/TOCF	1.44	1.48
CODT/TOCT	0.95	1.12
CODF/TOCF	1.20	1.16
	Wuhrmann's Experiments	
BODT/TOCT	0.99	1.48

Nitrogen and Phosphorus

Municipal wastewater contains other constituents in addition to BOD and suspended solids which may affect the design and performance of biological treatment plants. Specifically nitrogen and phosphorus are considered to be nutrients and indeed are required by the bacteria and other microorganisms responsible for the removal and degradation of the organic constituents of wastewater. Nitrogen and phosphorus are generally present in municipal wastewater at concentrations much higher than that required to sustain microbial growth. The excess nutrients are discharged in the effluent of a conventional biological treatment plant. The nitrogen and phosphorus along with carbon have been responsible for accelerating eutrophication or the aging process of streams and lakes.

The compositions of municipal wastewaters are presented in Table 1-4. The nitrogen and phosphorus concentrations illustrate the relative concentrations in comparison to the other components. Owens (1953) reported that the per capita phosphorus contribution to municipal wastewater varied from 1.5 to 3.7 grams per person per day with an average of about 2.0 gm/cap-day.

Table 1-4

Composition of Untreated Municipal Wastewater

	Culp (1967)	Merrell (1965)	Oswald (1961)
BOD mg/l	200-400	164-630	168
Phosphorus mg/l as P	8.3-10	12-32*	10.7
Nitrogen			
Organic mg/l as N	10-15	17-42	26.6
Ammonia mg/l as N	25-35	29-36	33.3
Nitrate and Nitrite mg/l as N	0	0.10-0.27	1.4
pH	7.2-7.4	6.9-7.8	9.3

*Orthophosphate

BASIC RELATIONSHIPS FOR BIOLOGICAL WASTEWATER TREATMENT

BIODEGRADATION OF ORGANIC COMPOUNDS

Municipal wastewater is composed of a mixture of dissolved, colloidal and particulate organic and inorganic materials. The concentration of any individual component is continuously changing as a result of sedimentation, hydrolysis, and microbial transformation and degradation of organic compounds.

The environment in a reactor permits the bacteria to use the organic material as a substrate for growth and a source of energy. When the substrate is depleted, the bacteria use cellular material for energy and complete auto-oxidation results in cell death. The organic material associated with these dead cells includes soluble biodegradable and particulate nonbiodegradable components.

The heterotrophic microorganisms which use organic compounds as a source of energy and synthesis play a most important role in the biological treatment of wastewaters. Autotrophic bacteria which oxidize inorganic compounds are also present. Typical autotrophic bacteria are the Nitrosomonas and Nitrobacter which oxidize ammonia to nitrite and nitrate nitrogen respectively. Algae, which are microscopic plants, are capable of autotrophic and heterotrophic synthesis. The role of algae will be discussed in connection with stabilization ponds.

Biological processes may be operated aerobically or anaerobically. In the first case free dissolved oxygen is maintained in the environment and microorganisms use this source of oxygen during the degradation of organic material. A simplification of this reaction is illustrated in Equation 2-1:

$$\text{Organic material} + O_2 + NH_3 + P \xrightarrow{\text{bacteria}} \underset{[(C_5H_7NO_2)_{10}P]}{\text{New Cells}} + CO_2 + H_2O \qquad \text{2-1}$$

The degradation of cell material is illustrated in Equation 2-2:

$$(C_5H_7NO_2)_{10}P + O_2 \longrightarrow CO_2 + H_2O + NH_3 + \text{Polysaccharide-like material} \qquad \text{2-2}$$

Free dissolved oxygen is not present in the anaerobic environment. In fact, oxygen is toxic to the truly anaerobic organisms. Under anaerobic conditions facultative bacteria convert organic material to intermediate organic acids and new cell material as illustrated in Equation 2-3.

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$$\text{Organic Matter} \xrightarrow[\text{Bacteria}]{\text{Facultative}} \text{Organic Acids + Cells} \qquad 2\text{-}3$$

Methane-forming bacteria which are true anaerobes convert the organic acids to methane and carbon dioxide gases as shown in Equation 2-4.

$$\text{Organic Acids} \xrightarrow[\text{Bacteria}]{\text{Methane}} CH_4 + CO_2 + \text{Cells} \qquad 2\text{-}4$$

The energy available to bacteria for synthesis and growth in an aerobic system is greater than under anaerobic conditions. The methane generated anaerobically contains most of the energy which otherwise would have been available for cell synthesis. Therefore, the quantity of sludge produced in anaerobic systems is much less than that generated during the aerobic biodegradation of organic material.

Biological utilization of organic material involves a series of enzyme catalyzed reactions. Simple dissolved organic material is easily incorporated in the cells and utilized for energy. However, the more complex organic materials must be dissolved or hydrolyzed by exoenzymes released outside the cell. This phenomenon may provide an explanation for the rapid removal of soluble organic material in the contact tank while the particulate and colloidal material undergo biodegradation in the stabilization tank of the contact stabilization process.

The aerobic biological treatment processes currently used in practice include various modifications of activated sludge process, trickling filters and aerated lagoons. Waste stabilization ponds are facultative systems in which aerobic and anaerobic zones of activity occur. The anaerobic process is generally applied for the digestion of sludges in many municipal systems.

The design and effective control of biological treatment processes require a basic understanding of the interdependence of the various biodegradation reactions. The basic correlations were developed from continuous flow reactors and reported by Herbert (1956), Monod (1949), and Schulze (1964). Glucose was the single substrate used in all these studies. Under these conditions only one type of microorganism will develop rather than the mixed culture which results from the use of the heterogeneous substrate such as municipal wastewater.

The general results of these studies are presented in Figure 2-1. These relationships indicate:

(a) The growth rate increases sharply at low concentrations of substrate; however, as the substrate concentration reaches a maximum a maximum rate of growth is reached.

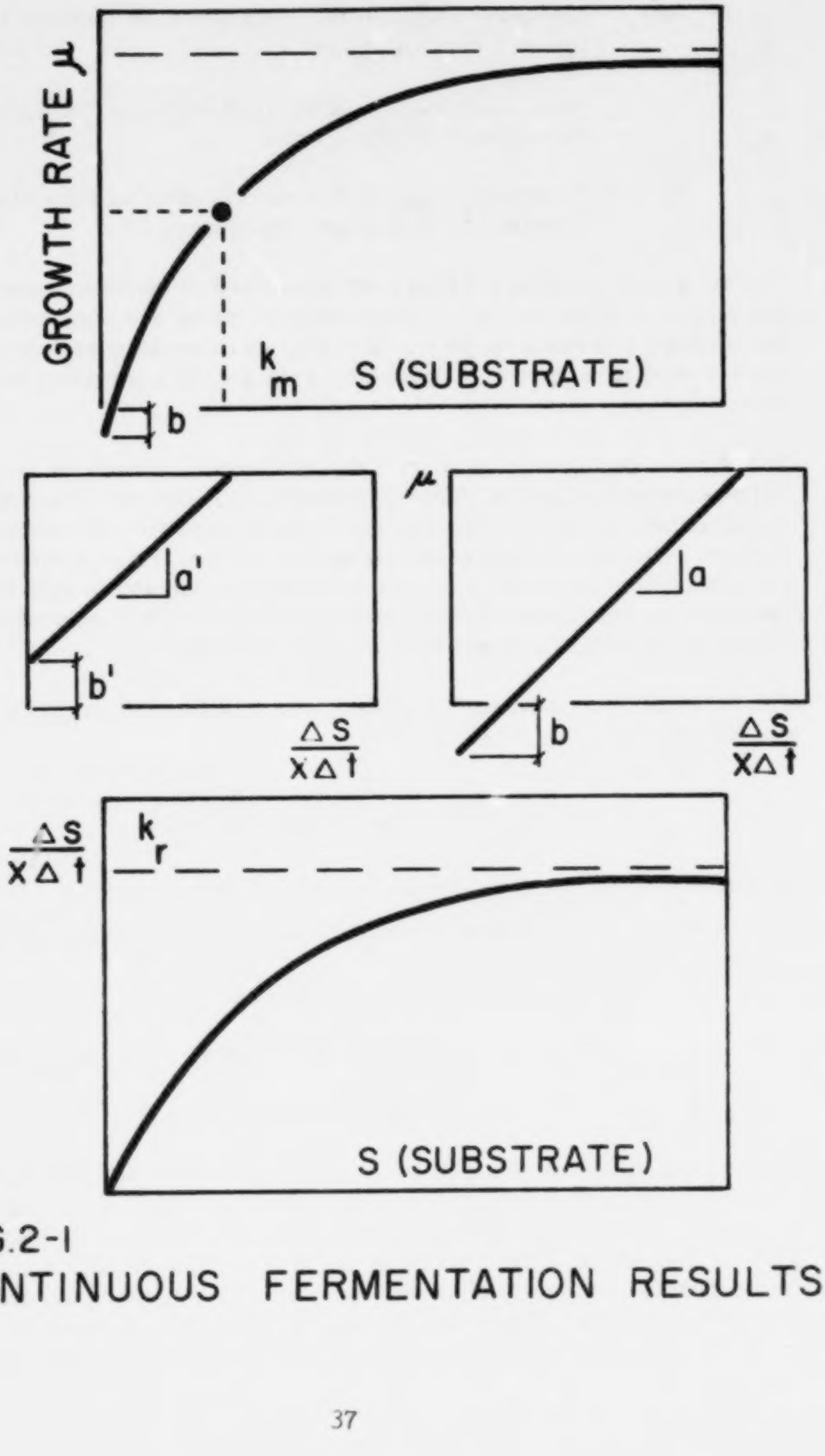

FIG.2-1
CONTINUOUS FERMENTATION RESULTS

(b) the oxygen uptake rate increases linearly as the substrate removal rate increases.

(c) there is a linear relationship between the growth rate and the substrate removal rate.

(d) substrate removal rate can be used to substitute for the growth rate in design procedures.

The substrate removal rate may be expressed in terms of pounds of BOD per pound of MLSS per day. High removal rates are possible only when the soluble substrate is high. However, at low substrate concentrations, the removal rate increases markedly with a slight increase in the substrate concentration.

These relationships are not directly applicable to multiple component substrates and mixed cultures of bacteria. Therefore, the mathematical formulations developed for glucose must be modified for municipal wastewaters. The equations presented below provide a basis for understanding the interaction between bacteria and a single substrate and are used to develop the applicable design equations for the various processes employed in the biological treatment of municipal wastewaters.

The relationship between the growth rate and the substrate concentration can be written in Equation 2-5.

$$\mu = \mu_{max} \frac{S}{k_m + S} - b \qquad 2\text{-}5$$

in which:

μ, μ_{max} = growth rate and maximum growth rate (day^{-1})

S = substrate concentration (mg/l)

k_m = substrate concentration at $\mu = 0.5\mu_{max}$ (mg/l)

b = endogenous respiration rate (day^{-1})

This equation was developed by Michaelis-Menten for enzyme kinetics and was first applied to describe growth kinetics in biological waste treatment by Monod (1959). The endogenous respiration rate, b, is frequently omitted from this equation. This endogenous rate is only detectable at extremely low substrate concentrations and compared to the maximum growth rate is almost negligible. However, the endogenous rate (b) must be considered in developing any growth kinetics or any substrate removal kinetics.

Theissier (1936) presented a different expression for the growth rate which is illustrated as Equation 2-6.

$$\mu = \mu_{max} (1 - e^{-cS}) - b \qquad 2\text{-}6$$

These equations are quite similar; however, the equation presented by Michaelis-Menten will be used to develop other design formulations. The relationships between oxygen uptake rate and growth rate with substrate removal rate are represented in Equations 2-7 and 2-8.

$$k_r = a' \frac{\Delta s}{X \Delta t} + b' \text{ (oxygen uptake rate)} \qquad 2\text{-}7$$

$$\mu = \mu \frac{\Delta X}{X \Delta t} = a \frac{\Delta S}{X \Delta t} - b \text{ (growth rate)} \qquad 2\text{-}8$$

in which:

k_r = oxygen uptake rate (mg O_2/mg cells-day)

$\frac{\Delta S}{X \Delta t}$ = substrate removal rate (mg substrate/mg cells-day)

a, a' = substrate utilization constants

b, b' = endogenous respiration constants (day^{-1})

X = concentration of active cells (mg/l)

Combining Equations 2-5 and 2-8, the substrate removal rate can be defined by Equation 2-9.

$$\frac{\Delta S}{X \Delta t} = k_R \left(\frac{S}{k_m + S} \right) \qquad 2\text{-}9$$

in which:

$k_R = \mu_{max}/a$ = maximum biodegradation rate (mg of substrate per mg of cells per day)

This equation is applicable only to completely mixed reactors. The relationship can be modified for plug flow or batch reactors as shown in Equation 2-10.

$$-\frac{dS}{Xdt} = k_R \left(\frac{S}{k_m + S}\right) \qquad 2\text{-}10$$

The negative sign indicates that the substrate removal rate decreases with increasing time. The integration of Equation 2-10 results in Equation 2-11.

$$\frac{S}{So} = e^{-\left(\frac{k_R}{k_m} Xt + \frac{S_o - S}{k_m}\right)} \qquad 2\text{-}11$$

The exponents in Equation 2-11 indicate the substrate removal is a function of time-dependent and substrate-dependent variables.

REMOVAL KINETICS

The expression presented in Equation 2-11 can be modified to describe the specific physical system used for waste treatment. The removal rate is constant if a single organic compound in a high concentration is to be treated. Under these conditions k_m is very small and the amount of substrate removed (S_o - S) is high; therefore, the first exponent is negligible and the substrate removal controls. However, linear removal is almost never observed for municipal wastewater. In general, it may be assumed that k_m is much larger than the effluent substrate concentration, S, and that (k_m + S) is proportional to k_m. Equation 2-9 can then be rewritten as Equation 2-12.

$$\frac{\Delta S}{X\Delta t} = kS \qquad 2\text{-}12$$

in which:

$$k = \frac{k_R}{k_m}$$

Integration of this expression results in Equation 2-13.

$$\frac{S}{S_o} = e^{-kXt} \qquad 2\text{-}13$$

The change in substrate concentration in a completely-mixed reactor can be described by a mass balance assuming that no biodegradation takes place, by Equation 2-14.

$$Q S_o \Delta t - Q S_e \Delta t - V\Delta S = 0 \qquad 2\text{-}14$$

This equation may be rearranged and presented as Equation 2-15.

$$\frac{\Delta S}{\Delta t} = \frac{Q (S_o - S_e)}{V} \qquad 2\text{-}15$$

Under steady-state conditions the change in substrate concentration with time is equal to the amount of substrate that must be removed; therefore, combining Equations 2-12 and 2-15 results in Equation 2-16.

$$\frac{Q (S_o - S_e)}{V} = kXS \qquad 2\text{-}16$$

The hydraulic detention time is equal to the volume divided by the rate of flow $(t = \frac{V}{Q})$, and in a completely-mixed reactor the effluent concentration S_e is equal to the concentration of substrate in the reactor (S); therefore, the effluent concentration may be defined by Equation 2-17.

$$S = \frac{S_o}{1 + kXt} \qquad 2\text{-}17$$

These equations deal primarily with the soluble material and must be modified for municipal wastewaters since only 40 to 60 percent of the total BOD entering the system is soluble. A relatively short period of time of mixing the incoming wastewater with the activated sludge is necessary for the removal of the soluble components of the waste. The rate of removal is dependent on the mixed liquor suspended solids concentration. The kinetics of removal of soluble organic compounds was described in Equation 2-12. The effect of aeration time on the substrate removal rate and the quantity of substrate remaining are illustrated in Figure 2-2. A higher removal rate is observed at short aeration times; however, as the aeration time increases, the amount of BOD remaining decreases and the overall substrate removal rate decreases until an equilibrium level is reached. Beyond this equilibrium point the effluent BOD concentration remains constant and the removal rate continues to decrease for a period.

The concentration of microorganisms in the activated sludge process can be controlled to some extent by the rate of sludge wasting. However, in the aerated lagoon process the level of microorganisms is self-controlled and is affected by the overall detention time. The rate of growth of microorganisms in the activated sludge process was presented in equation 2-8 as a ratio of the weight of sludge produced during the process per weight of sludge in the reactor, $(\frac{\Delta X}{X\Delta t})$. The inverse of the growth rate is in fact the

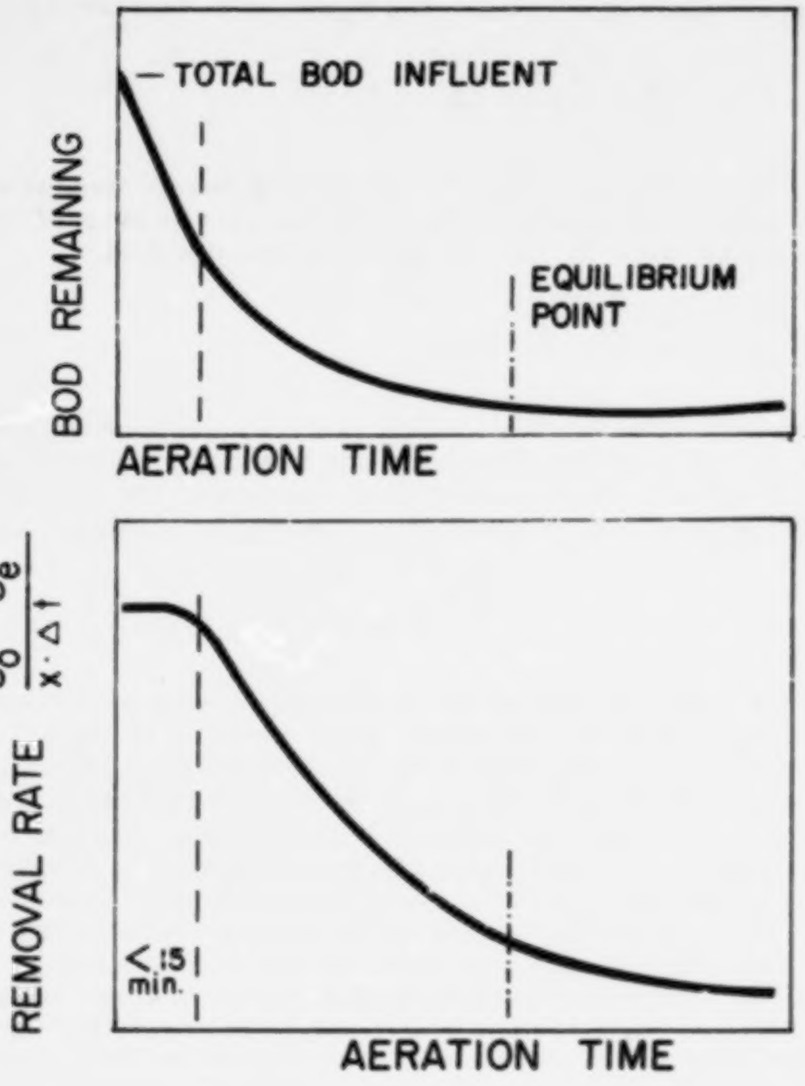

FIG.2-2
REMOVAL RATE AND REMAINING
BOD AS FUNCTIONS OF AERATION TIME

sludge age (G) of the solids in the activated sludge process which represents the length of time the sludge is in the system. The sludge age is equal to the hydraulic detention time in flow through processes such as the aerated lagoon process. Equation 2-8 can be rewritten in terms of sludge age by using Equation 2-12 to define the substrate removal rate as Equation 2-18. This equation can be further reduced to Equation 2-19 which can be rearranged and the following expression for effluent substrate concentration results.

$$\frac{1}{G} = \frac{1}{t} = a\,\frac{S}{X\Delta t} - b \qquad 2\text{-}18$$

$$\frac{1}{t} = akS - b \qquad 2\text{-}19$$

$$S = \frac{1}{akt} + \frac{b}{ak} \qquad 2\text{-}20$$

This equation indicates that the effluent soluble substrate concentration in a flow-through system such as the aerated lagoon and certain waste stabilization ponds is a function of only the detention time and is independent of influent concentration. The suspended mixed liquor solids or algae which develop in the process are not included in the substrate. The effects of the influent and effluent substrate concentrations and the required detention time can be calculated by a trial and error approach or by a graphical solution using Equations 2-21 and 2-22.

$$X_a = \frac{a\,(S_o - S)}{1 + bt} \qquad 2\text{-}21$$

$$X_a = \frac{(S_o - S)}{Skt} \qquad 2\text{-}22$$

in which X_a indicates the active microorganism population in the process. The curves presented in Figure 2-3 are based on an experimentally determined constant, k=0.20, a=0.65, and b = 0.15 and represent the interrelationship among the active mixed liquor suspended solids, the remaining BOD, the detention time and the incoming BOD for an aerated lagoon system. The data indicate that:

(a) The mass of active mixed liquor suspended solids is dependent on the concentration of the BOD in the incoming wastewater.

(b) There is a maximum concentration of active mixed liquor suspended solids which occurs at a one-day detention time for influent BOD concentrations from 50 - 100 mg/l.

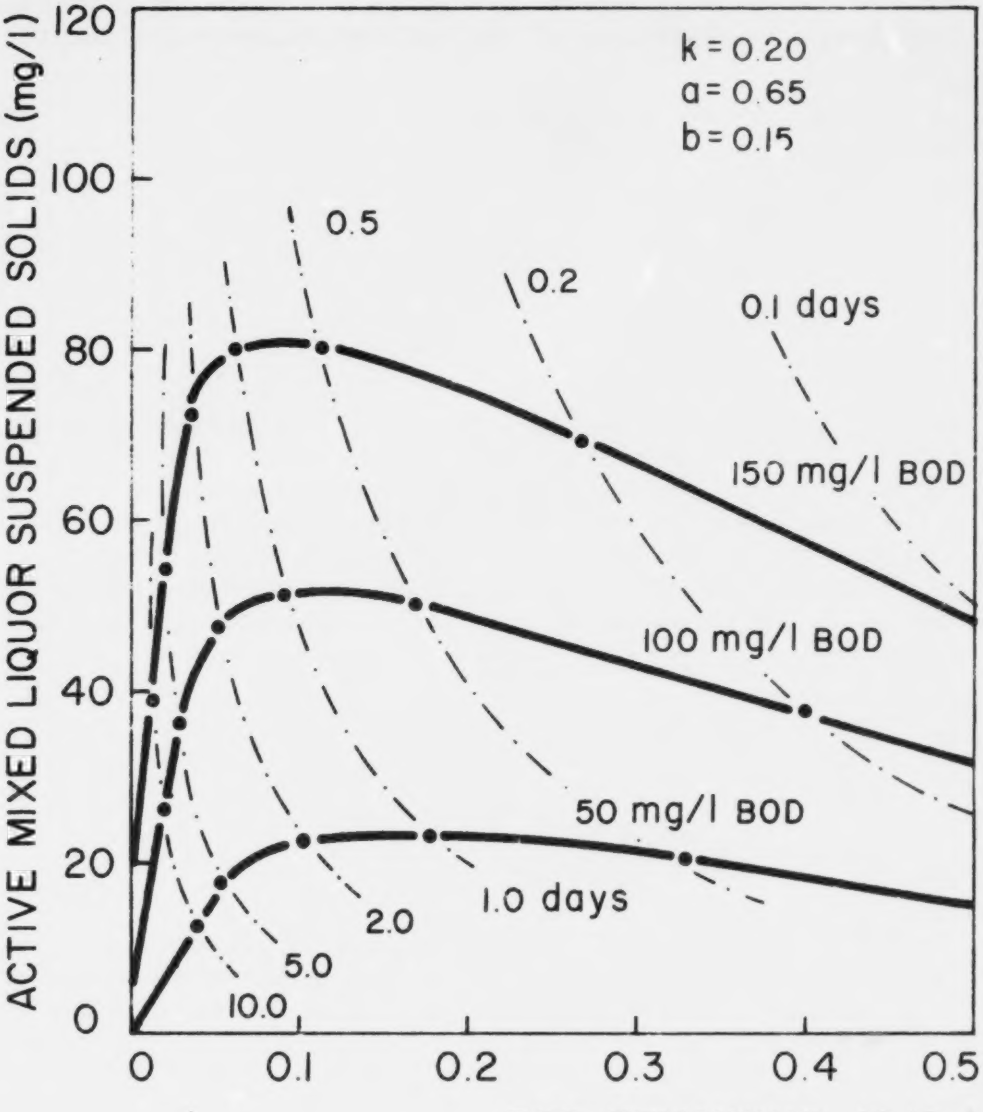

FIG.2-3

AERATED LAGOON INTERACTIONS

(c) At the one-day detention time, the concentration of active mixed liquor suspended solids increases with the increasing concentration of incoming BOD.

(d) The percent BOD remaining increases from about ten percent to about 20 percent as the BOD coming into the system decreases from 150 to 50 mg/l at a one-day detention time.

(e) The concentration of active mixed liquor suspended solids decreases as the detention time increases above one day. This decrease in suspended solids is a result of endogenous respiration or autooxidation of the synthesized cellular material.

(f) The active mixed liquor suspended solids concentration is also lower at detention times less than one day. This phenomenon may be attributed to the fact that sufficient time is not available for synthesis of the incoming BOD to cellular material. At the shorter detention times the removal of BOD is only in the range of from 50 to 80 percent. However, the concentration of active mixed liquor suspended solids is not markedly different at the half-day detention time than at the one-day detention time.

(g) The maximum removal of BOD occurs at the maximum active mixed liquor suspended solids concentration and increases as the detention time increases beyond that time at which the maximum active mixed liquor suspended solids concentration develops. At the maximum mixed liquor suspended solids concentration, the removal of BOD is in excess of 90 percent.

The maximum active mixed liquor suspended solids in aerated lagoons, is important when considering the kinetics of substrate removal. The data indicate that at detention times longer than that required for development of maximum mixed liquor suspended solids concentration, and in this case more than one day,the rate of substrate removal decreases. The results of full-scale experimentation with aerated lagoons at detention times between three and eight days indicate that the soluble BOD in the effluent was about the same in all cases. The detention time at which the maximum active mixed liquor suspended solids concentration occurs is similar to the equilibrium point illustrated in Figure 2-2. At higher concentrations of mixed liquor suspended solids in the activated sludge process, the equilibrium point is reached at shorter detention times than in the aerated lagoon process. However, it could be expected that the equilibrium point in the activated sludge process and in the aerated lagoon process occur at about the same sludge age.

Smith and Eilers (1969) reported that the data observed at the Hyperion Treatment Plant indicate a decrease in the rate constant with decreasing load on the activated sludge process and with increasing sludge age. The sludge age in these experiments ranged from one to eight days and at various sludge ages the soluble effluent BOD was relatively constant.

The removal of soluble organic material by biological treatment is controlled to a great extent by the condition of the mixed liquor suspended solids. The activated sludge solids must settle readily and concentrate to a minimum volume to reduce the return sludge pumping requirements. The results of various studies indicate that the effluent suspended solids concentration and the total effluent BOD decreased with increased aeration time and sludge age. Therefore, most biological treatment plants normally are operated at some detention time beyond the equilibrium point and the reaction rate constant which is a function of the average organic loading to the system, fluctuates with the organic loading.

The overall performance of a biological treatment system involves flocculation, adsorption, and settling in addition to biological degradation, stabilization ponds and aerated lagoons. The effluents contain suspended solids, algae, and bacteria which represent a major portion of the organic material leaving the process. Data observed for the biological treatment of municipal wastewater may be correlated using Equation 2-17. It should be pointed out that the observed rate constants represent all the phenomena by which BOD is removed and are more applicable for design purposes.

EXCESS SLUDGE

The quantity of excess sludge produced during biological treatment may be expressed by the modified growth model presented as Equation 2-23.

$$\frac{\Delta X_v}{X_v t} = a \frac{S_o - S_e}{X_v t} - b \tag{2-23}$$

The constants, a and b, vary widely and are affected to a large extent by the influent concentration of suspended solids. Eckenfelder (1967) reported values of a = 0.73 and b = 0.075 per day for the activated sludge process treating settled municipal wastewaters. However, Garrett (1958) reported values of a = 1.10 and b = 0.08 per day for activated sludge treating raw municipal wastewater. The microbial solids generated in the process must be distinguished from the nonbiodegradable suspended solids. The total sludge mass generated and leaving the system can be calculated from Equation 2-24.

$$\frac{WX_{RS} + QX_e}{X_aV} = \frac{\Delta X_a}{Xt} + \frac{\Delta X_n}{Xt} \qquad 2\text{-}24$$

in which:

W = waste sludge flow (MGD)

X_{RS} = concentration of suspended solids in return sludge (mg/l)

Q = wastewater flow (MGD)

X_e = concentration of effluent suspended solids (mg/l)

X_a = MLSS concentration (mg/l)

X = Mass of solids in aeration tank (lb)

V = Volume of aeration tank (million gallons)

t = aeration time (days)

ΔXa = active biological solids formed (lb)

ΔXn = nonbiodegradable suspended solids accumulated (lb)

therefore:

WX_{RS} = sludge wasted daily (lb/day)

QX_e = sludge lost daily in effluent (lb/day)

X_aV = sludge mass in aeration tank (lb)

$\frac{\Delta X_a}{Xt}$ = rate of biological solids formation (lb/lb - day)

$\frac{\Delta X_n}{Xt}$ = rate of accumulation of nonbiodegradable solids (lb/lb - day)

The sludge age (G) in days may therefore be expressed as:

$$G = \frac{X_aV}{WX_{RS} + QX_e} \qquad 2\text{-}25$$

and Equation 2-24 may be transformed to:

$$\frac{1}{G} = \frac{\Delta X_a}{Xt} + \frac{\Delta X_n}{Xt} \qquad 2\text{-}26$$

These mathematical models do not include the nondegraded biodegradable suspended solids since at the range of sludge ages commonly used in practice, the quantity of these solids is approximately constant and are included in the nonbiodegradable fraction. The rate of formation of biological solids can be determined from Equation 2-27 which includes the total suspended solids concentration in the mixed liquor.

$$\frac{\Delta X_a}{Xt} = a^* \frac{(S_o - S_e^*)}{Xt} - b^* \qquad 2\text{-}27$$

in which:

a^* = synthesis of soluble BOD removed to suspended solids

b^* = rate of endogenous respiration of active solids (lb/lb - day)

$\frac{(S_o - S_e^*)}{Xt}$ = rate of BOD removal $\left(\frac{\text{lb BOD}}{\text{lb MLSS - day}}\right)$

S_e^* = filtered effluent BOD

The excess sludge includes the influent nonbiodegradable suspended solids and the rate of accumulation may be expressed as:

$$\frac{\Delta X_n}{Xt} = \frac{QX_{on}}{XV} = \frac{X_{on}}{Xt} \qquad 2\text{-}28$$

in which:

X_{on} = nonbiodegradable suspended solids in the influent (mg/l)

Equation 2-26 may be rewritten as:

$$\frac{1}{G} = a^* \frac{(S_o - S^*_e)}{Xt} - b^* + \frac{X_{on}}{Xt} \qquad 2\text{-}29$$

In this equation, the effluent concentration is expressed as the BOD of the filtered sample.

Typical values of the constants in Equation 2-29 are $a^* = 0.6$ to 0.65 pounds of active cells formed per pound of BOD removed, $b^* = 0.075$ pound per pound per day and the nonbiodegradable suspended solids in the influent are equal to 60 percent of the total solids in the influent. These data may be used for preliminary calculations or for checking existing calculations if no other information is available. This model for excess sludge production should be applicable to all aerobic processes; however, there are some limitations. In low rate trickling filters and in stabilization ponds, algae and other microorganisms are produced and tend to control the quantity of solids which are generated. Therefore, Equation 2-29 is not applicable in these cases in the present form.

OXYGEN UPTAKE

The general equation for oxygen uptake rate was developed and presented as equation 2-7; however, this equation can be rewritten in a more practical form as Equation 2-30.

$$R = a' \frac{(S_o - S^*_e)}{t} + b'X \qquad 2\text{-}30$$

in which:

R = Oxygen uptake rate (lb/day)

$\frac{S_o - S^*_e}{t}$ = BOD removal rate (lb/day)

X = mass of mixed liquor suspended solids (lb)

a' = oxygen required for synthesis of BOD removed $\left(\frac{\text{lb Oxygen}}{\text{lb BOD}}\right)$

b' = oxygen required for endogenous respiration $\left(\frac{\text{lb Oxygen}}{\text{lb MLSS-day}}\right)$

Typical values of a' and b' for municipal wastewater are a'=0.53 (0.40 to 0.65) and b'=0.15 per day, respectively. The value for b' is affected markedly by the temperature;and the value presented is typical of summer conditions and is widely used. Equation 2-35 represents the oxygen required for the biodegradation of organic material; however, at low organic loading rates, nitrification can be expected. The oxygen requirements for nitrification may be expressed as a stoichiometric relationship developed by Downing (1964):

$$2\ NH_4 + 4\ O_2 \longrightarrow 2\ NO_3 + H_2O + 4\ H^+ \qquad 2\text{-}31$$

This reaction indicates that approximately 4.6 pounds of oxygen were required to convert one mole of ammonia-nitrogen to nitrate nitrogen. The oxygen per pound of nitrogen converted can be calculated as $\frac{64}{14} = 4.60$. The rate of oxygen uptake for nitrification can be expressed as Equation 2-32.

$$R_N = 4.60\ \frac{\Delta NH_3}{t} \qquad 2\text{-}32$$

in which:

R_N = oxygen uptake rate for nitrification (lb/day)

ΔNH_3 = ammonia nitrogen removed (lb)

t = aeration time(days)

In those treatment plants which have a low organic load to the aeration system, partial nitrification has been reported. Consequently, if nitrification is not included in calculations of oxygen required, higher values of the coefficient, a', must be used in Equation 2-30. The total oxygen including nitrification can be expressed as:

$$R = a'\ \left(\frac{S_o - S_e^*}{t}\right) + b'X + 4.60\frac{\Delta NH_3}{t} \qquad 2\text{-}33$$

DESIGN EQUATIONS AND APPLICATIONS

The various design equations for biological treatment processes are summarized in Table 2.1. The applicability of the various equations and limitations are also included in this table. Typical values of the constants used in the equations are presented.

TABLE 2-1

	Activated Sludge	Aerated Lagoons	Trickling Filter	Waste Stabilization Pond
$\frac{S}{S_o} = e^{-kXt}$	Batch tests plug flow system	------	Modified form	Ponds in series ajd. baffled lab units
$\frac{S}{S_o} = e^{-kAvDQ^{-n}}$	------	------	$S^*/S_o^* = 0.50$ $k^* = 0.010 - 0.020$ (Q=MGAD)	------
$\frac{S}{S_o} = \frac{1}{1 + kXt}$	Only high load rates G < 1 day	------	------	------
$\frac{S}{S_o} = \frac{1}{1 + kt}$	------	Detention time < 2 days k = 8-15	------	Overall performance $k \sim 0.1 - 0.2$
$S = \frac{1}{akt} + \frac{b}{ak}$	------	only t < 2 days a=0.63, b=0.15 k=0.10-0.20	------	------
$R = a'\left(\frac{S_o - S_e^*}{t}\right) + b'X_v$	$a' = 0.53$ $b' = 0.15$	$a' = 0.53$, $b' = 0.15$, Completely mixed lagoons only	not needed for design	------
$R = a'\left(\frac{S_o - S_e^*}{t}\right) + b'X_v + 4.60\,\frac{\Delta NH_3}{t}$	$a' = 0.53$ $b' = 0.15$	$a' = 0.53$, $b' = 0.15$, Completely mixed lagoons only	not needed for design	------
$\frac{1}{G} = a\left(\frac{S_o - S_e^*}{X_v t}\right) - b$	a = 0.75 (settled) a = 1.10 (raw) b = 0.075	------	in a modified form	------
$\frac{1}{G} = a^*\left(\frac{S_o - S_e^*}{X_v t}\right) - b^* + \frac{X_{on}}{Xt}$	$a^* = 0.63$; $b^* = 0.075$ $X_{on} \approx 0.6\,X_o$	------	in a modified form	------
$\frac{1}{G} = a^*\left(\frac{S_o + X_o}{Xt}\right) - b^*$	$a^* = 0.6$ $b^* = 0.075$	------	------	------
$t_T = t_{20}\,\theta^{20 - T}$ $k_T = k_{20}\theta^{20 - T}$	$\theta = 1.0 - 1.04$	$\theta = 1.03 - 1.10$	$\theta = 1.035$	$\theta = 1.085$

MIXING MODELS FOR REACTOR

Reactors used in wastewater treatment are aeration tanks, aeration lagoons, trickling filters, waste stabilization ponds and anaerobic digestion tanks. Mixing is essential to the maintenance of a high rate and a high degree of biodegradation. In the activated sludge process and the aerated lagoon process, mixing is accomplished by aeration which also provides the oxygen essential to the function of the process. Waste stabilization ponds are mixed by the action of wind along the surface which tends to turn over the entire body of water. The microorganisms in the trickling filter are fixed as slime layers on the surface of the packing medium. Therefore, the mixing is accomplished by distributing the incoming wastewater uniformly over the entire surface of the filter medium.

The kinetic models presented earlier in this chapter describe either of two idealized flow systems, namely the completely mixed or the plug flow system. In the completely mixed system, the incoming material is instantaneously distributed throughout the entire tank volume; therefore, the concentration in the tank is the same at all locations. Typical flow patterns for the two types of systems are illustrated in Figure 2-4. Theoretically, no longitudinal mixing takes place in the plug flow system, and any material introduced with the influent will be associated with this liquid parcel as it passes through the tank. An aeration tank in which the waste is introduced at one end and travels longitudinally to the effluent at the other end and trickling filters may be considered plug flow reactors.

Longitudinal plug flow-type tanks were originally preferred for the activated sludge process because early investigators indicated that the performance of the system was primarily based on the aeration or contact time of the biological solids in the mixed liquor and the substrate in the wastewater. In the plug flow system, the aeration time was predictable; however, the possibility of shortcircuiting was expressed by Hurd (1929), Calvert and Bloodgood (1934), Haseltine (1932), Kessener (1935), and Kehr (1936). The more recent literature indicates that extensive studies in which tracer response techniques were applied in order to determine the "residence time distribution" of fluid elements in the reactor have been published.

Several models have been proposed to characterize the non-ideal flow patterns in aeration tanks, namely:

(a) subdivided tank model
(b) tanks-in-series model
(c) dispersion model

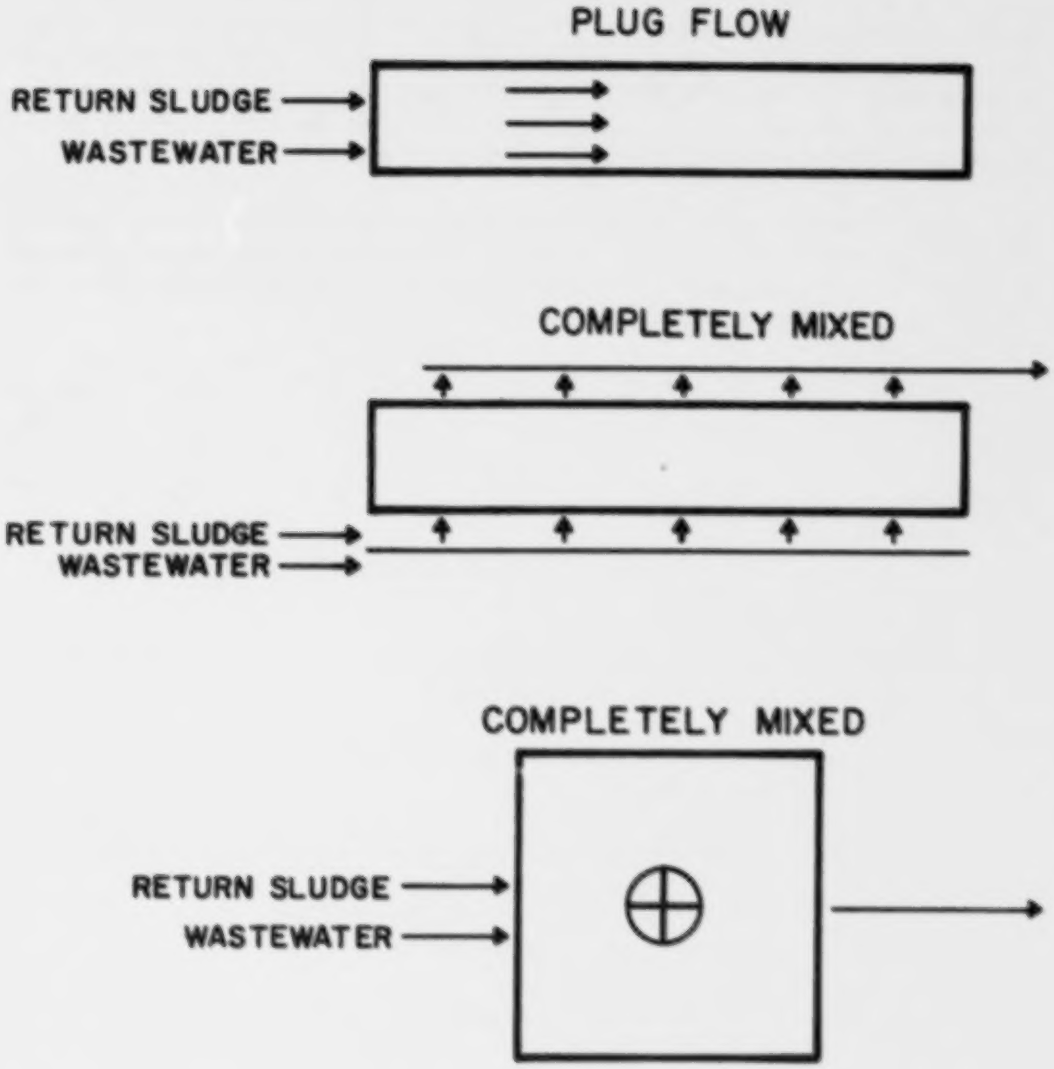
PLUG FLOW
RETURN SLUDGE
WASTEWATER
COMPLETELY MIXED
RETURN SLUDGE
WASTEWATER
COMPLETELY MIXED
RETURN SLUDGE
WASTEWATER

FIG.2-4 AERATION TANK TYPES

The subdivided tank model describes the conditions which exist, if the original tank was subdivided into smaller individual flow regions connected in series and/or in parallel. Each of the individual regions is usually restricted to the simple case of plug flow, completely mixed, single parameter dispersion, and dead space. The type of flow connecting the various regions can be categorized as cross-flow, bypassing flow, and recycle flow from end to beginning of the regions. These generalized models have a large number of parameters which must be evaluated and in practice are difficult to determine. Levenspiel (1962) reported that "an unrealistic many parameter model may closely fit all present data after the fact, but may be quite unrealistic for prediction in new untried situations."

The tank-in-series model is somewhat simpler to understand. An aeration tank is visualized as a number of equal-volume completely mixed compartments. The effluent concentration of a tracer for this particular model can be calculated by the following equation.

$$\frac{C}{C_o} = \frac{n^n}{(n-1)!} \left(\frac{t}{\bar{t}}\right)^{n-1} e^{-nt/\bar{t}} \qquad 2\text{-}34$$

in which:

t = time measured from the initial addition of the tracer (hours)

$\bar{t}$ = the theoretical detention time = the volume of the tank divided by the flow (hours)

C = tracer concentration at the exit of the tank measured at time t (mg/l)

C_o = average tracer concentration = mass of tracer added divided by the tank volume (mg/l)

n = number of tanks in series

This model describes the mixing in an aeration tank in terms of one single parameter, n, the number of completely mixed tanks in series. The value of n may be determined from experimental response curves by comparing dimensionless response curves with a set of curves with various values of n.

Levenspiel (1962) related the number of tanks to the variance α^2 of the experimental response curve as shown in Equation 2-35.

$$n = \frac{1}{\sigma^2} \qquad 2\text{-}35$$

The number of tanks calculated from Equation 2-35 will not in general be an integer. This paradox cannot be theoretically justified because in equation 2-34, n is assumed to be an integer.

Thomas and McKee (1944) reported good correlation between experimental and theoretical response curves which were predicted by Equation 2-34 for a compartmentalized laboratory aeration tank. This model was also used for the evaluation of data observed during mixing studies in aerated lagoons at the Williamson Creek Wastewater Treatment Plant in Austin, Texas. These data are presented in Figure 2-5. The recovery of the tracer in the effluent (C/C_o) is plotted versus the ratio detention time $t/\bar{t}$. The data indicate that at energy inputs of 0.03 HP/1000 gallons of tank volume the tracer recovery is very close to the theoretical recovery curve. The initial peak seems to be the result of some short circuiting. The results observed at a power level of 0.015 HP/1000 gallons indicate a deviation of the actual tracer recovery curve from that expected for an ideal completely mixed basin. Therefore, at the lower power level a number of completely mixed tanks-in-series would better represent the situation. The theoretical number of tanks-in-series was 1.42, 1.2, and 4.1, respectively at power levels of 0.054, 0.032, and 0.015 HP/1000 gallons. The actual value of the numbers of tanks-in-series should not be overemphasized since a repetition of the test could result in different distribution and the actual value may be higher or lower than a previous test. These results do indicate that at the higher power levels, this particular tank will function more like a completely mixed system than at lower power levels.

The dispersion model was considered to describe longitudinal mixing in reactors by Dankwerts (1953), Levenspiel (1957, 1959, 1962) and Van der Laan (1958). An analogy between mixing in actual flow and molecular diffusion was presented; therefore, back mixing of fluid flowing in the axial direction may be represented by an equation similar to Fick's second law of molecular diffusion. Equation 2-36 represents the mathematical model considering a uniform intensity of backmixing.

$$\frac{\partial C}{\partial t} = D\frac{\partial^2 C}{\partial x^2} \qquad 2\text{-}36$$

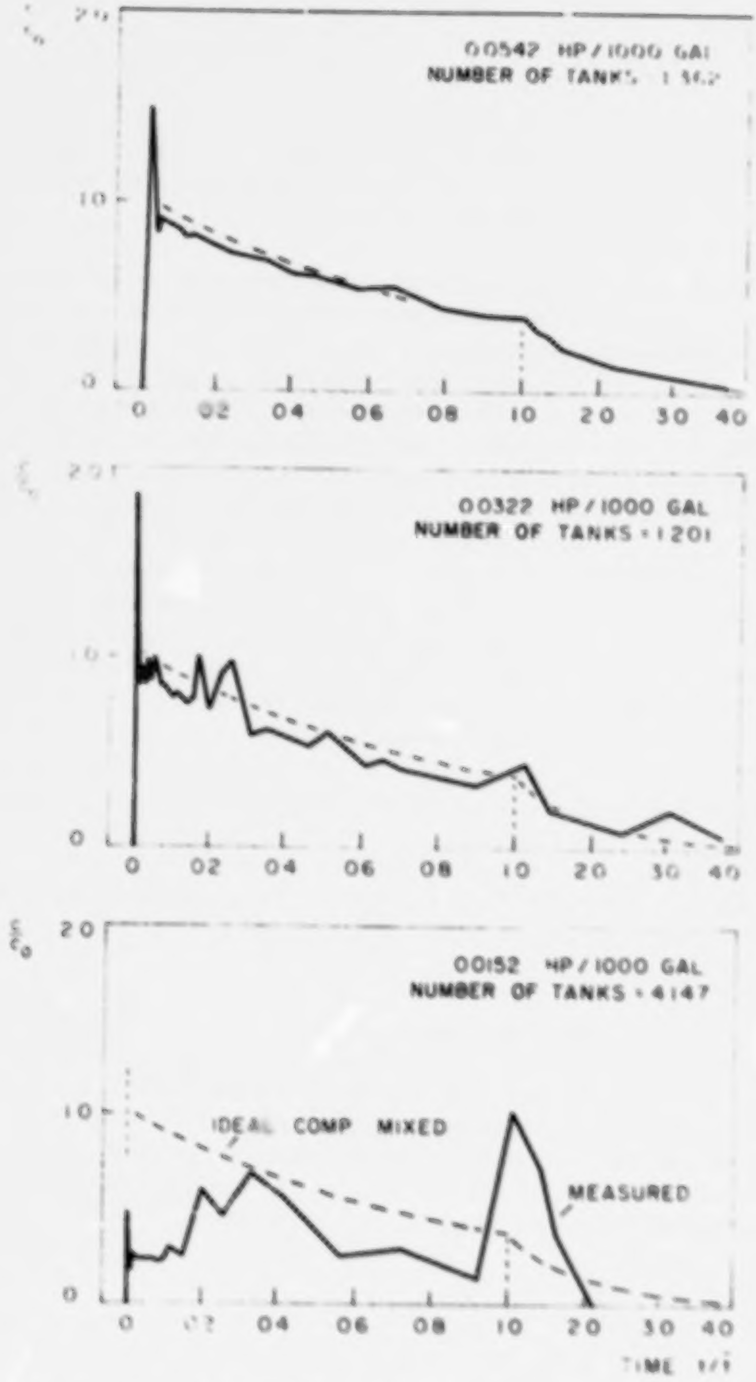

FIG. 2-5 RESULTS OF TRACER STUDIES FROM AERATED LAGOONS

in which:

C = the concentration

t = time

x = longitudinal distance

D = longitudinal dispersion coefficient which characterizes the degree of backmixing

The application of this model has been previously limited to flow through tubular and packed bed reactors. The dispersion term includes molecular and turbulent or eddy diffusion. The term longitudinal is used in order to distinguish between radial mixing or dispersion. The application of the dispersion model assumes that lateral mixing is sufficient to insure a uniform concentration of tracer at any given cross-section. Solutions to the second order differential equation for a tracer pulse input to a closed vessel were developed by Thomas and McKee (1944) and Miyauchi (1953). A dispersion number was developed as a constant in the solutions of these equations. This dispersion number is presented in Equation 2-37.

$$\frac{D}{\mu L} \qquad 2\text{-}37$$

in which:

D = dispersion coefficient (sq ft/hr)

μ = bulk velocity along the tank length = flow rate of water divided by tank cross section (ft/hr)

L = tank length (ft)

Van der Laan (1958) and Timpany (1966) developed techniques for estimating dispersion number from experimental response curves. The dispersion coefficients can be derived from this dispersion number.

Boyko (1968), Boyko and Murphy (1968) and Timpany and Murphy (1967) reported the results of measurements on laboratory and full-scale spiral flow aeration tanks and the conclusions are summarized as follows:

(a) The dispersion coefficient is dependent on the geometry of the tank cross section and the specific air flow rate. The following expression may be used to calculate the dispersion coefficient.

$$D = 3.118\, W^2\, q_A^{\,0.346} \qquad \text{2-38}$$

in which:

W = tank width (ft)

q_A = specific air flow rate (scfm/1000 cu ft of tank volume)

The dispersion coefficients measured for full-scale tanks ranged from 5000 to 7500 sq ft/hr.

(b) The dispersion coefficient was not affected by water temperature in the range of 12° to 30°C.

(c) There was no significant difference in the value of the dispersion coefficient observed in experiments using coarse bubble aeration (sparger) and fine bubble aeration (saran wrapped tubes) at the same air flow rates.

The dispersion coefficient also represents the degree of backmixing in a particular cross-secion of the tank. The dispersion number, however, is dependent on the dispersion coefficient, detention time, and tank length. The dispersion number can be expressed in terms of the average time:

$$\frac{D}{uL} = \frac{D\bar{t}}{L^2} \qquad \text{2-39}$$

Substituting the general expression for the dispersion coefficient, the dispersion number can be redefined as:

$$\frac{D}{uL} = 3.118 \left(\frac{W}{L}\right)^2 \bar{t}\, q_A^{\,0.346} \qquad \text{2-40}$$

The constants in Equation 2-40 must be verified experimentally; however, the terms 3.118 and 0.346 are presented at this point to indicate the relative degree of influence that each constant has on the dispersion number. Therefore, air flow rate has a relatively small influence on the mixing in a spiral flow tank compared to other variables. The tank geometry on the other hand has the most influence on the mixing. Shallow wide tanks with long detention times more closely approximate the completely mixed system in which the dispersion number ($\frac{D}{uL}$) approaches infinity. On the other hand, long narrow tanks with relatively short detention times more closely approximate plug flow conditions in which the dispersion number approaches zero. The results of tracer studies in one aeration tank at the Govalle Wastewater Treatment Plant in Austin, Texas indicate that dispersion coefficient was equal to about 3500 sq ft/hr (Halbert and Malina, 1970). This value of dispersion coefficient is much lower than those reported by Murphy, Boyko, and Timpany (1967, 1968).

The method of evaluating the data from tracer studies has considerable influence on the results obtained. The aeration tank at the Govalle Plant is not a single straight longitudinal tank but is divided in the middle so that the flow pattern is around the end. The transition in flow pattern occurring at the mid-point of the tank may have an effect of the dispersion coefficient. The degree of backmixing at the opening between the two tanks may have been less than expected in the normal tank cross-section. The cross-sectional areas of the tank and the opening were 375 square feet and 150 square feet, respectively. The measured dispersion number was 0.06 sq ft/hr and indicates that the tank represents a plug flow system.

The measured response curves for around the end tanks is shown in Figure 2-6. Each curve represents the response curves at different stations located along the length of the tank. Station 6 was located in the effluent of the tank and this response curve indicates that actual detention time was 0.86 of the theoretical detention time. Station 1 was located at a distance equal to a volume which represents about 40 percent of the theoretical detention time.

Batch biological reactors were also operated during the tracer studies in order to determine a biodegradation rate constant. Theoretically the quantity of substrate remaining along the length of the tank is predictable, if the biodegradation rate constant and the dispersion number are known. Dankwerts (1953) developed an equation to incorporate these two parameters and a solution for this particular equation was developed, Wehener and Wilhelm (1957). This equation was used to evaluate the rate constant data and the dispersion number; however, the measured soluble substrate concentration along the length of the tank decreases at a much faster rate than the calculated value (Halbert and Malina, 1970).

Additional mixing studies should be carried out in tanks of different geometry and with different diffuser arrangements in order to determine the condition of flow. In aerated lagoons and tanks operated at low power levels, short-circuiting can easily be detected by tracer studies. These tracer studies should be conducted at conditions as close to steady state operation as

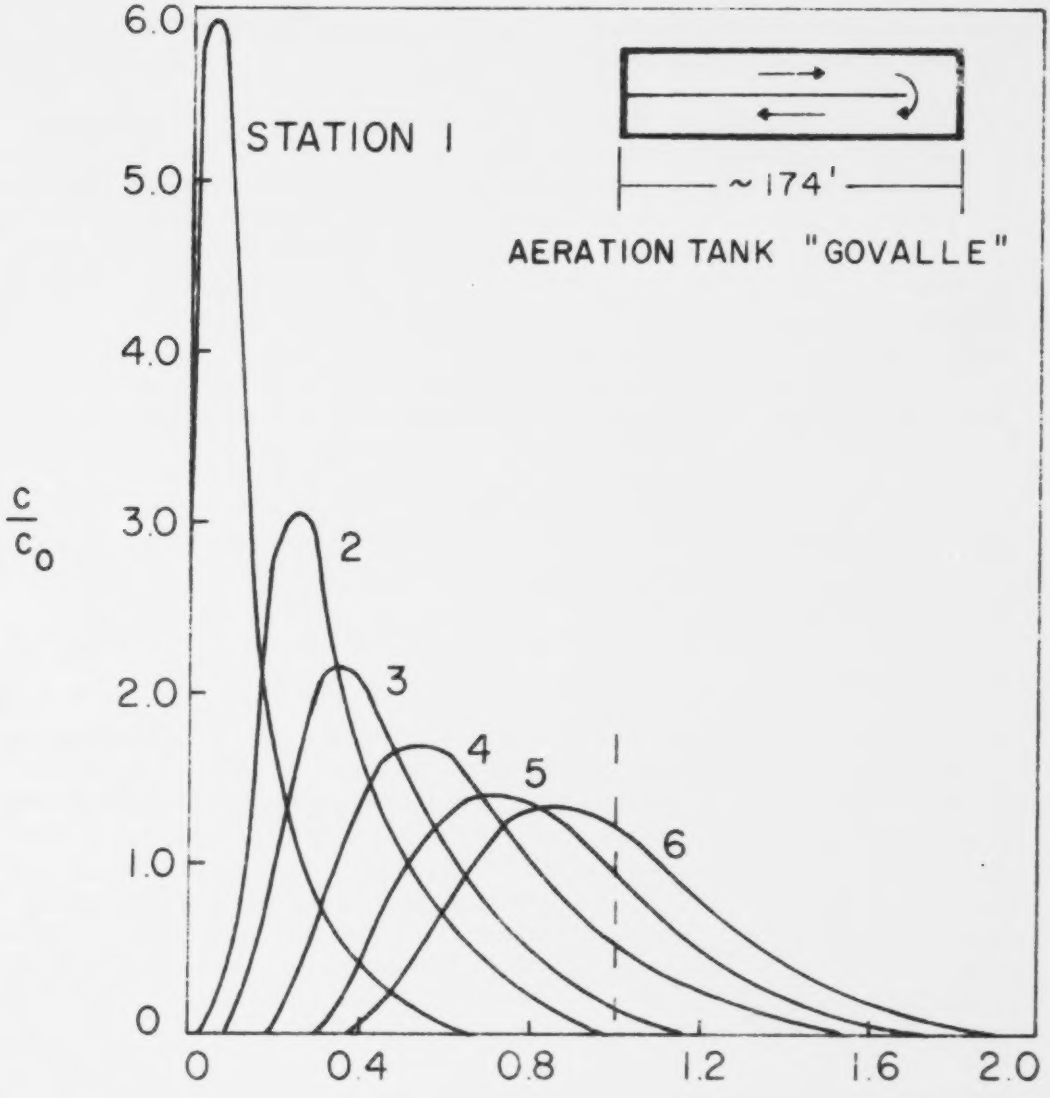

FIG. 2-6
TRACER RESPONSE CURVES

possible. Development of dissolved oxygen profiles, substrate profiles and determination of rate constants should accompany the tracer studies. In aerated lagoons the tracer study data should be supplemented by profiles of suspended solids at various locations and at different depths in the lagoons.

ACTIVATED SLUDGE PROCESS

PROCESS DESCRIPTION

The activated sludge process is probably the most commonly used system for the biological treatment of municipal and industrial wastewaters. The essential components of the activated sludge plant are an aeration tank with the necessary aeration equipment, a final clarifier in which the biological solids can be separated from the liquid effluent, and the necessary pumps to recycle the concentrated solids to the aeration tank. The incoming wastewater is mixed with the returned sludge solids in the aeration tank and this mixture is aerated for about two to six hours. The intensity of aeration must be sufficient to mix the tank contents. The mixture of incoming wastewater and the activated sludge solids is commonly called the mixed liquor. Most of the settled sludge is returned from the clarifier to the aeration tank; however, a portion of the sludge is wasted. This excess sludge is a result of bacterial growth as well as the accumulation of nonbiodegradable suspended solids which enter the system.

The activated sludge contains a mixture of a wide variety of microorganisms which are capable of degrading different organic compounds. Inactive suspended materials also accumulate in the biological floc which forms during the aeration. The bacteria and other microorganisms in the activated sludge are generally found in most municipal wastewaters. The mass of activated sludge increases as the microorganisms degrade the organic material in the wastewater and grow in the aerobic environment of the aeration tank. The bacterial growth rate during the initial stages is very high because the substrate concentration is relatively high compared to the bacterial population. After a week or two of operation of the activated sludge treatment plant, the desired mixed liquor suspended solids concentration and the desired degree of treatment are attained.

The activated sludge process has been modified in numerous ways since 1914 when the process was first applied to wastewater treatment. In the original process which is called the conventional activated sludge process, the wastewater and return sludge entered at one end of a long narrow tank. The mixed liquor flowed in the longitudinal direction and with the help of the diffused air aeration system, a spiral flow resulted. The mixed liquor left the other end of the long narrow tank.

A number of the modifications in the activated sludge process are illustrated in Figure 3-1. The basic differences occur in the point at which the incoming

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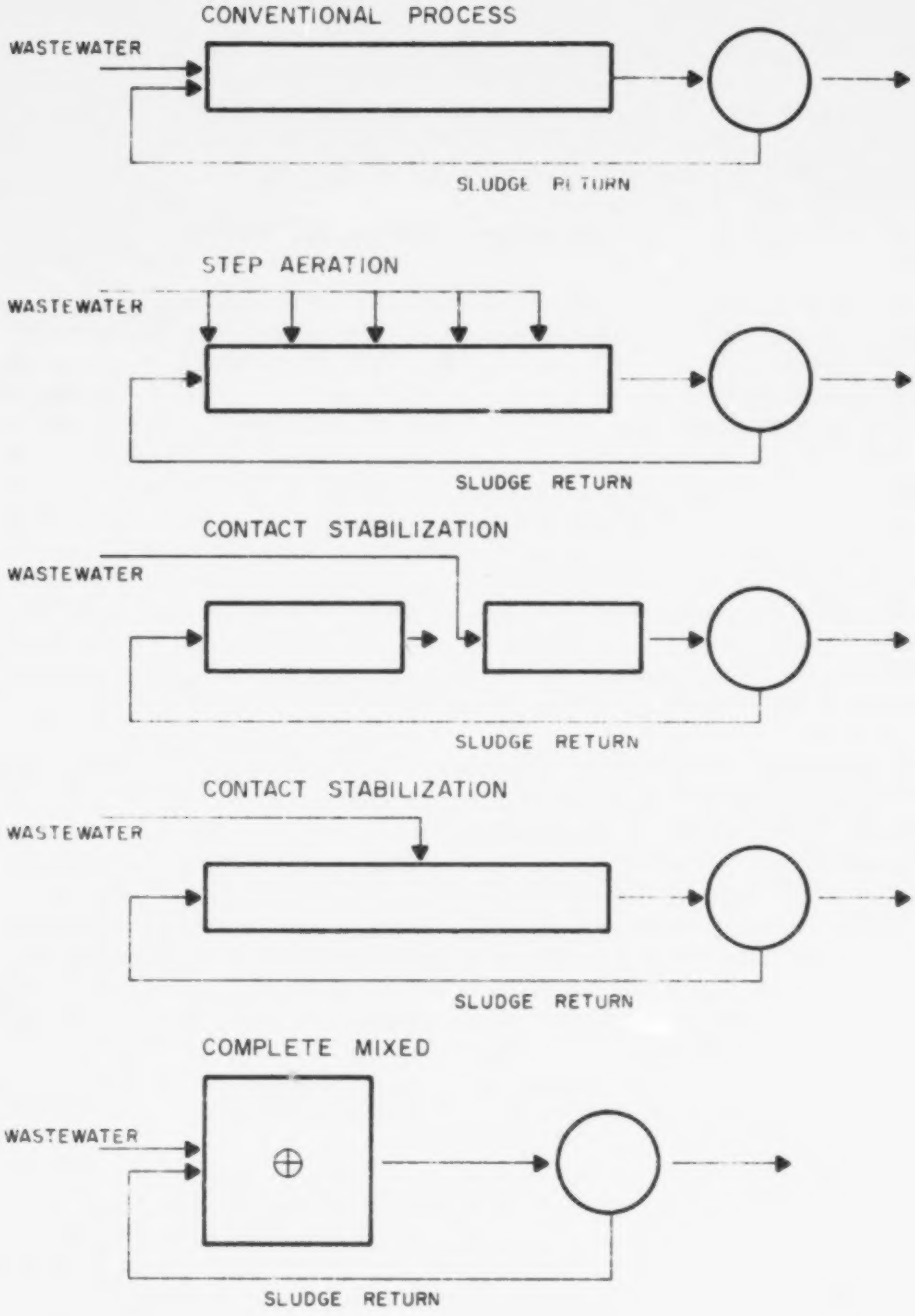

CONVENTIONAL PROCESS
WASTEWATER
SLUDGE RETURN
STEP AERATION
WASTEWATER
SLUDGE RETURN
CONTACT STABILIZATION
WASTEWATER
SLUDGE RETURN
CONTACT STABILIZATION
WASTEWATER
SLUDGE RETURN
COMPLETE MIXED
WASTEWATER
SLUDGE RETURN

FIG. 3-1 AERATION TANK ARRANGEMENTS

wastewater is introduced into the aeration tank and mixed with the return sludge, the reaeration of the return sludge prior to coming into contact with the incoming wastewater and the type of flow pattern that might be typical of the system and operation.

Aeration of the return sludge has been practiced for many years. The original purpose for reaeration was to maintain the return sludge aerobic because the sludge would turn anaerobic in the final clarifier since the quantity of return sludge was relatively small. Kraus (1945) and Hatfield (1931) pumped anaerobic digester supernatant into the reaeration tank and effectively treated the supernatant with the return sludge. In both cases the concentration of mixed liquor suspended solids was high and a long aeration time was maintained. This type of system also minimized the shock loads to the aeration tank since the flow of digester supernatant was distributed more uniformly throughout the day and a higher concentration of solids was available to utilize the organic material in the supernatant.

The contact stabilization process was introduced independently by Ullrich and Smith (1951) as well as by Eckenfelder and Grich (1956). This process was based on observations made in the conventional activated sludge system which indicated that the soluble fraction of the organic material in wastewater was almost entirely removed after very short aeration times. Therefore in the contact stabilization process, the incoming wastewater is mixed with the return reaerated sludge for very short times of about 30 minutes to two hours. The sludge solids are then separated from the liquid and the concentrated sludge is reaerated prior to being mixed with the incoming wastewater. The aeration times in the contact zone must be long enough to permit a reduction of the concentration of soluble organics to the predetermined effluent concentration. However, in the contact zone, the nonsoluble colloidal and particulate materials are also incorporated in the activated sludge floc particles and are removed in the final clarifier. The sludge is aerated for an additional two to four hours in the stabilization or reaeration tank. In many plants, the stabilization zone and the contact zone are not physically separated; therefore, it is difficult to calculate the exact time which may be attributed to contact and to stabilization.

The completely mixed system has been used extensively in laboratory-scale units in evaluating the various parameters which affect the activated sludge process. However, this process has only been recently installed for the treatment of municipal and industrial wastewaters. The development of surface and mechanical aerators which may be used separately or in conjunction with diffused air systems has made the use of completely mixed tanks more popular. In the completely mixed tanks, the concentration of oxygen and of the substrate is constant at each point throughout the entire tank volume. The system permits automatic control of oxygen transfer and is relatively resistant to shock loading since the influent wastewater is distributed uniformly throughout the entire tank volume.

The extended aeration process is a modification of the activated sludge process and is characterized by relatively low organic loadings; therefore, the sludge which is produced during biodegradation is also stabilized to some extent in the aeration tank. However, in the extended aeration process the nonbiodegradable solids tend to accumulate and must be removed periodically.

The activated sludge process can be operated as the conventional activated sludge process and the contact stabilization process. The flow pattern in the aeration tanks can be categorized as the plug flow or the completely mixed. The performance of the various process modifications and flow schemes can only be compared if each system is operated at an identical organic loading and with identical wastewater. The results of laboratory-scale experiments comparing contact stabilization and the conventional activated sludge processes indicated that the effluent BOD concentration for each process was almost identical at organic loading rates of 0.40 to 1.90 pounds of BOD/pound of mixed liquor suspended solids -day (Water Pollution Research Laboratory, 1967, 1968).

PERFORMANCE OF THE ACTIVATED SLUDGE PROCESS

The first order substrate removal model does not normally apply for the organic loadings at which most activated sludge plants are operated. Therefore, data are compared on the basis of detention time, BOD volume load (pounds of BOD per 1000 gallons of tank volume per day) and organic loading (pounds of BOD per pound mixed liquor suspended solids or per pound of mixed liquor volatile suspended solids per day). However, neither the MLSS nor MLVSS concentrations define the inert fraction of the solids present in the mixed liquor. The nonbiodegradable solids in the mixed liquor include the inert suspended solids in the influent and in the return sludge and may be represented as Equation 3-1 which can be modified and used to calculate the nonbiodegradable suspended solids.

$$QX_{on} = WX_n + QX_{en} \qquad 3\text{-}1$$

$$X_n = X_{on} \cdot \frac{G}{t} \qquad 3\text{-}2$$

in which:

X_{on}, X_{en} = influent and effluent nonbiodegradable suspended solids (mg/l)

X_n = nonbiodegradable suspended solids in the aeration tank (mg/l)

C = sludge age (days) (total suspended solids in the system divided by the total suspended solids leaving the system per day)

t = theoretical detention time (days) (aeration tank volume divided by wastewater flow rate)

The interrelationship between the concentration of nonbiodegradable suspended solids in the mixed liquor, the sludge age, and the theoretical aeration time are shown in Figure 3-2. These curves are based on an influent nonbiodegradable suspended solids concentration of 50 mg/l. Therefore, at a sludge age of four days and a theoretical detention time of six hours the accumulation of nonbiodegradable suspended solids in the mixed liquor is about 750 mg/l.

The relationship of the nonbiodegradable fraction of the mixed liquor suspended solids and the ratio of the influent BOD to influent suspended solids is presented in Figure 3-3. These data are from pilot-scale studies conducted by Wuhrmann (1964) and from the experimental work conducted at the Govalle Treatment Plant in Austin, Texas, and the Hyperion Treatment Plant in Los Angeles, California. The accumulation of nonbiodegradable suspended solids in the mixed liquor range from 40 to 70 percent of the total mixed liquor suspended solids and was affected by the ratio of influent BOD to influent suspended solids (S_o/X_o) which ranged from 0.5 to 1.75.

The curves in Figure 3-3 reflect the efficiency of primary clarification at the particular treatment plant. The relatively high percentage of nonbiodegradable solids in the mixed liquor at the Govalle Plant is indicative of no primary clarification. The relatively low ratio reported for the Hyperion Plant is typical of efficient primary treatment. The data reported by Wuhrmann (1964) for the operation of a pilot plant reflect the different analytical procedures used for determining suspended solids. The Wuhrmann data are based on a membrane filtration technique which captures about 10 to 15 percent more of the suspended solids than the glass fiber or paper filter techniques used at the Govalle and Hyperion Plants. If the results reported by Wuhrmann are adjusted, the data would compare quite favorably.

The operating conditions under which these data were collected were influent BOD of 90 to 200 mg/l, and influent suspended solids concentrations of 70 to 250 mg/l. The mixed liquor suspended solids concentrations was 600 to 6000 mg/l and the BOD loading rate varied from 0.17 to 2.0 lb BOD/lb MLSS-day. The curves indicate that the influence of the loading is relatively

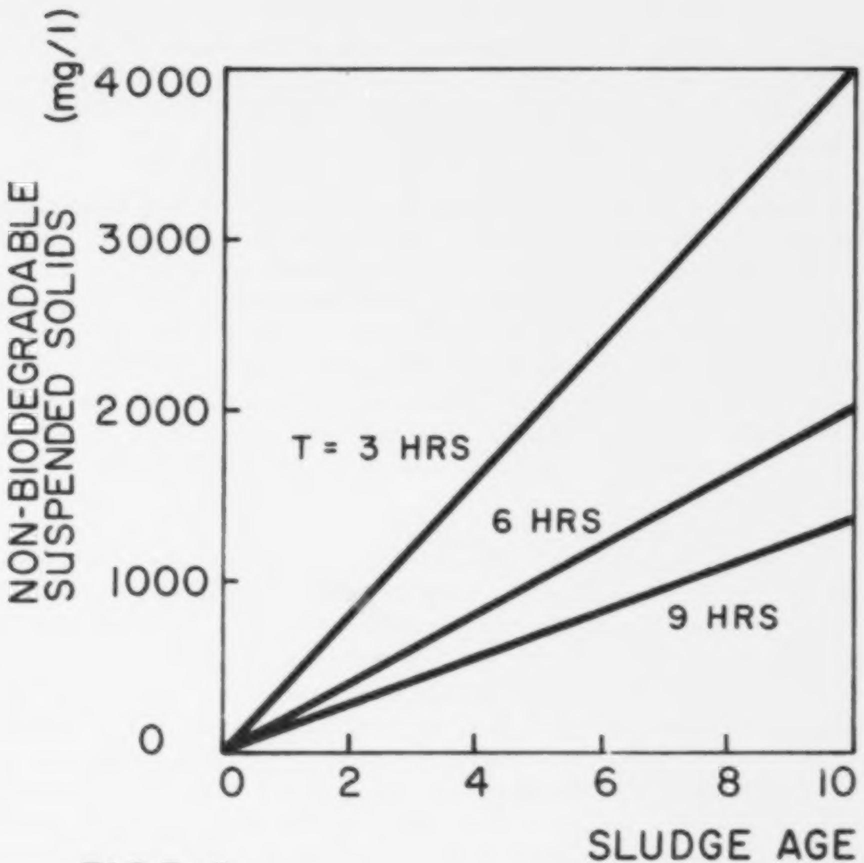


FIG.3-2
NON-BIODEGRADABLE SUSPENDED SOLIDS IN MIXED LIQUOR

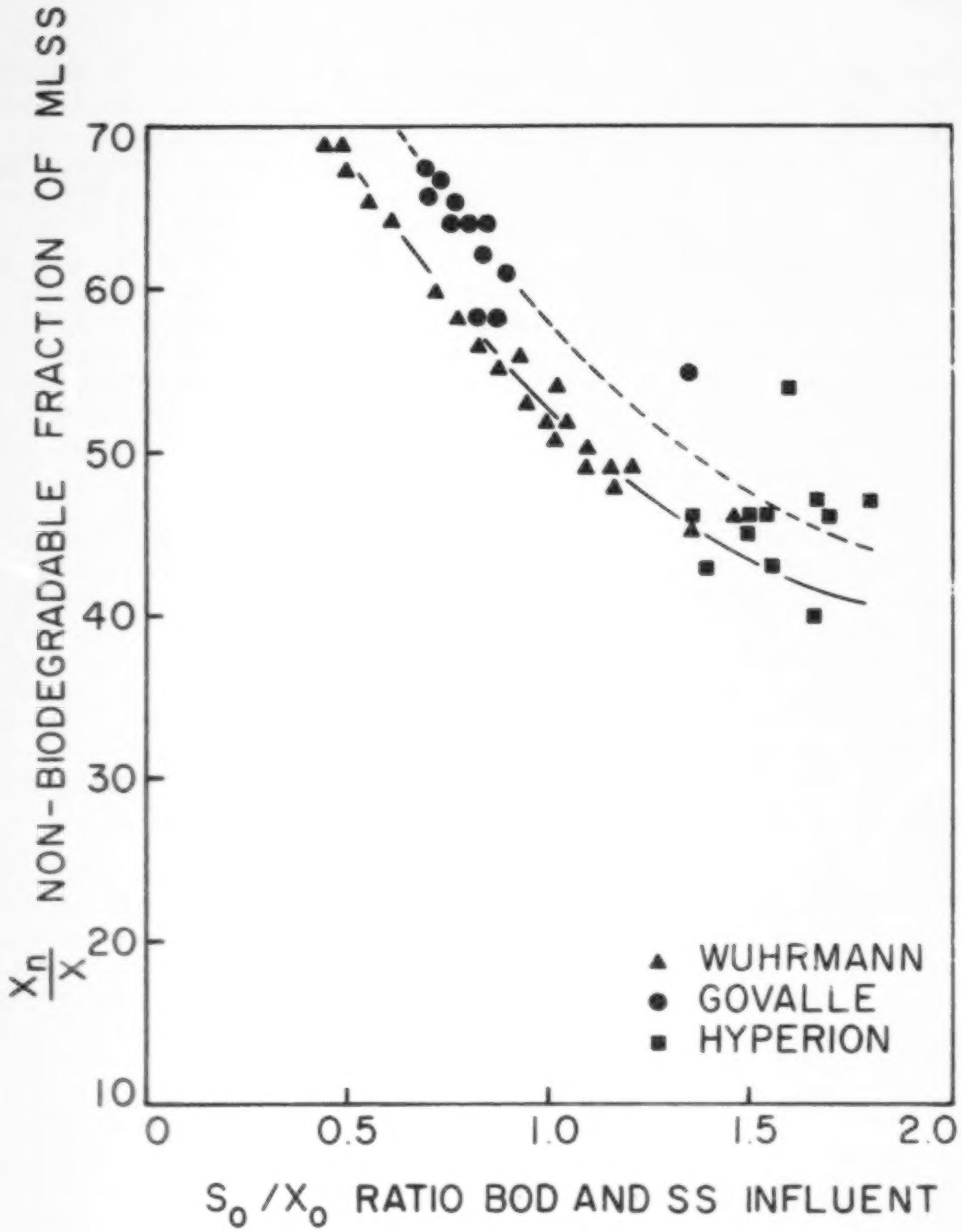


FIG.3-3

NON-BIODEGRADABLE SOLIDS OF MIXED LIQUOR negligible At times, it may be useful to compare data based on the mixed liquor active suspended solids concentration (MLASS) which can be obtained by subtracting the nonbiodegradable fraction from the total suspended solids concentration. For example, at a loading rate of 0.3 lb BOD/lb MLSS - day, the rates of BOD removal based on the active mixed liquor suspended solids concentration would be 0.55 and 0.85 lb of BOD/lb MLASS - day, respectively. The ratio of MLASS to MLSS for Hyperion and Govalle were 0.55 and 0.35, respectively. These data indicate that the loading rate at the Govalle Plant is about 60 percent higher at the Hyperion Plant. Loading based on volatile suspended solids would exhibit similar differences. However, the mixed liquor suspended solids (MLSS) concentration is generally the only measure of the solids in the aeration tank at most plants; therefore, a comparison of performance data for various plants must be based on the total mixed liquor suspended solids.

The relationship between the effluent filtered BOD and the BOD loading (lb BOD/lb MLSS - day) is illustrated in Figure 3-4. The BOD loading ranged from 0.2 to 1.0 lb BOD/lb MLSS - day. The effluent filtered BOD of the Govalle samples scattered widely from about three mg/l to over 11 mg/l for loadings of less than 0.2 to about 0.5 lb BOD/lb MLSS - day. The scatter in the data from these experiments was considerably greater than that for the data observed at Hyperion which indicated a slight increase in the effluent BOD as the loading increased from 0.2 to almost 1.4 lb BOD/lb MLSS - day. The Hyperion data seemed to indicate a linear relationship between the effluent filtered BOD and the BOD loading range.

It should be pointed out that data presented in Figure 3-4 are not comparable theoretically since the Hyperion Plant and one of the three plants at the Govalle Plant were operated as the conventional activated sludge process whereby the other two plants at the Govalle Plant were operated as the contact stabilization process. If one considers that the removal rate of soluble substrate is not affected by the nonsoluble organic material, the relationship which results comparing the soluble BOD loading rates and the effluent soluble BOD concentration can be calculated and plotted as shown in Figure 3-5. The curves indicate that there is a linear relationship between the effluent soluble BOD and the soluble BOD loading. The loading rates for the contact stabilization plant were based only on the volume of the contact zone, the MLSS in the contact zone, and the incoming wastewater flow. These data indicate that the concentration of soluble material in the effluent increases as the loading expressed as soluble BOD also increases. In general, at the same soluble BOD loading the residual BOD at the Govalle Plant was higher than that reported at the Hyperion Plant. The results of laboratory-scale experiments at the Govalle Plant indicate that the soluble BOD in the effluent of contact stabilization process was between three and six mg/l at soluble organic loading rates of 0.3 to 2.3 lb BOD/lb MLSS - day

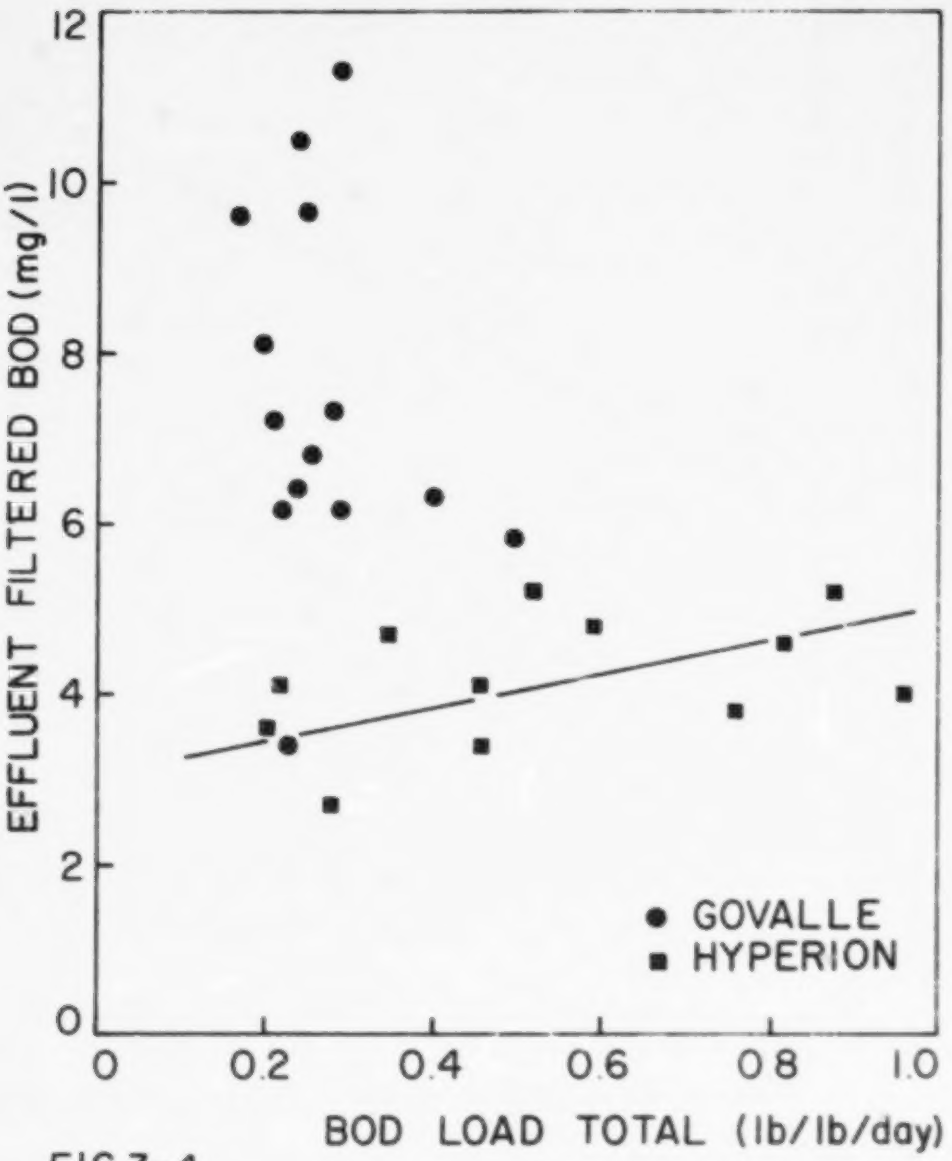


FIG.3-4

TOTAL LOAD AND SOLUBLE EFFLUENT

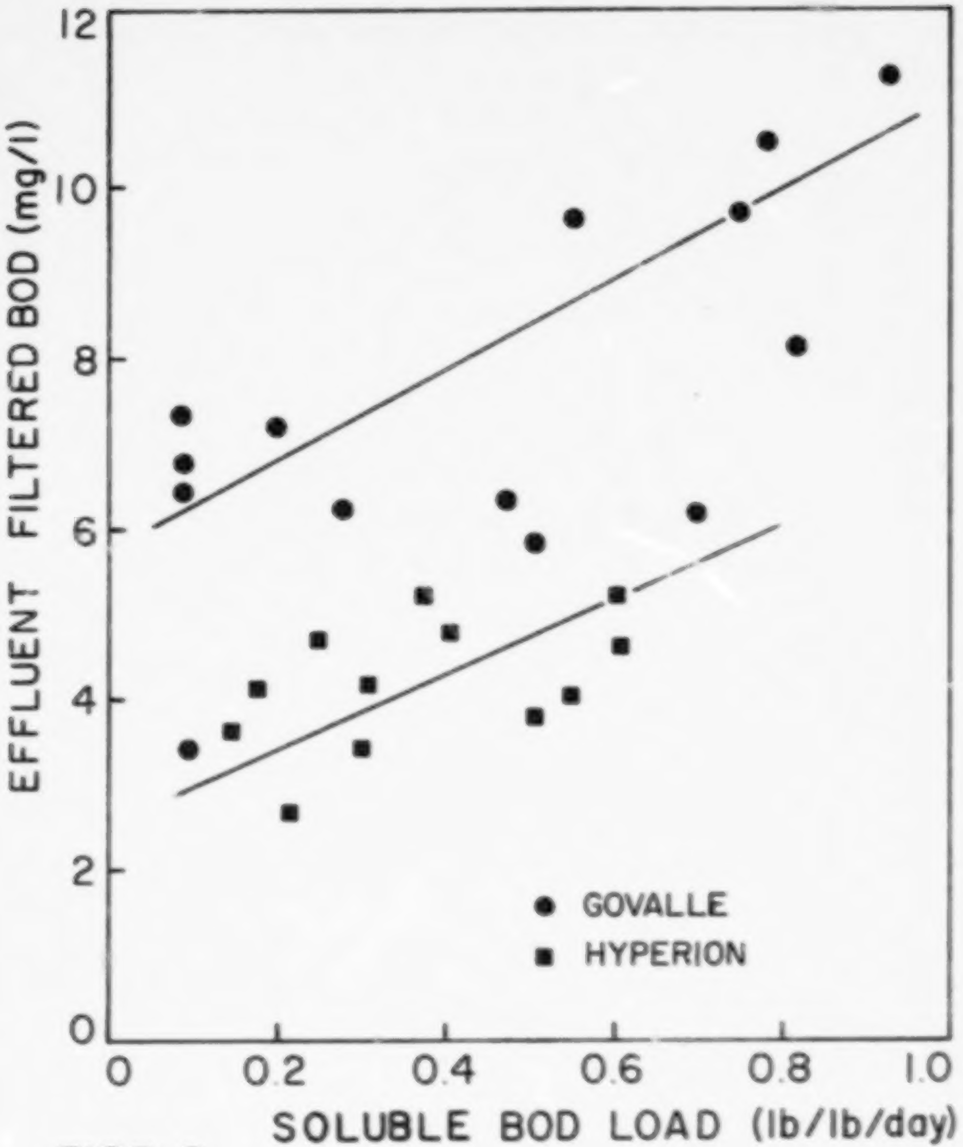


FIG.3-5

SOLUBLE LOAD AND EFFLUENT

(Berryhill, 1970). These residual BOD concentrations are in the same range as that reported for the Hyperion Plant effluent. The relationship between the effluent soluble COD and the soluble COD loading rates are presented in Figure 3-6. These data indicate that the effluent soluble COD increases at a much more rapid rate with increasing soluble COD loading at the Hyperion Plant than at the Govalle Plant.

The discrepancy between the laboratory-scale data and the full-scale data observed at the Govalle Plant may be explained by the fact that the laboratory-scale runs were performed about six months after the full-scale experiments, and the composition of the wastewater was different. The relatively high effluent BOD at the Govalle Plant indicate that these residual organic materials are degradable but at an extremely show rate.

The more useful presentation of data for practical purposes is the relationship between the total effluent BOD and the loading rate expressed in terms of total BOD entering the plant. This relationship is presented in Figure 3-7, for data observed at the Hyperion Plant, the Govalle Plant, and the pilot-plant studies by Wuhrmann (1964). These data scatter considerably over a loading rate of about 0.1 to almost 2.0 lb BOD/lb MLSS - day. The results of the three different experiments should be discussed individually.

The results for the Hyperion Plant were reported by Smith and Eilers (1969). The wastewater flow for all the experiments was 50 MGD. The theoretical aeration times were four, five, and six hours, respectively, and were maintained by using different numbers of aeration tanks. The loading rate was changed by varying the mixed liquor suspended solids concentration in the aeration tank from 600 to 3500 mg/l. The surface area of the settling tanks were constant for all experiments and provided an overflow rate of about 530 gallons/sq ft-day. The increase in total BOD in the effluent was the result of the increased effluent suspended solids concentration caused by the increased loading rates.

During the Govalle experiments, the wastewater flow varied slightly and the organic loading was changed by controlling the mixed liquor suspended solids concentration in the aeration tanks. The overflow rates for the clarifiers fluctuated from 750 to 1000 gallons/sq ft - day. Therefore, part of the scatter of the effluent BOD is caused by the suspended solids in the effluent. However, during most of the runs, partial nitrification took place. It is therefore safe to assume that the nitrification continued in the BOD bottle resulting in higher BOD concentrations than would be expected for only carbonaceous material.

Wuhrmann on the other hand, used three pilot plants in parallel in the experiments in Zurich, Switzerland. The detention times used were 25, 50,

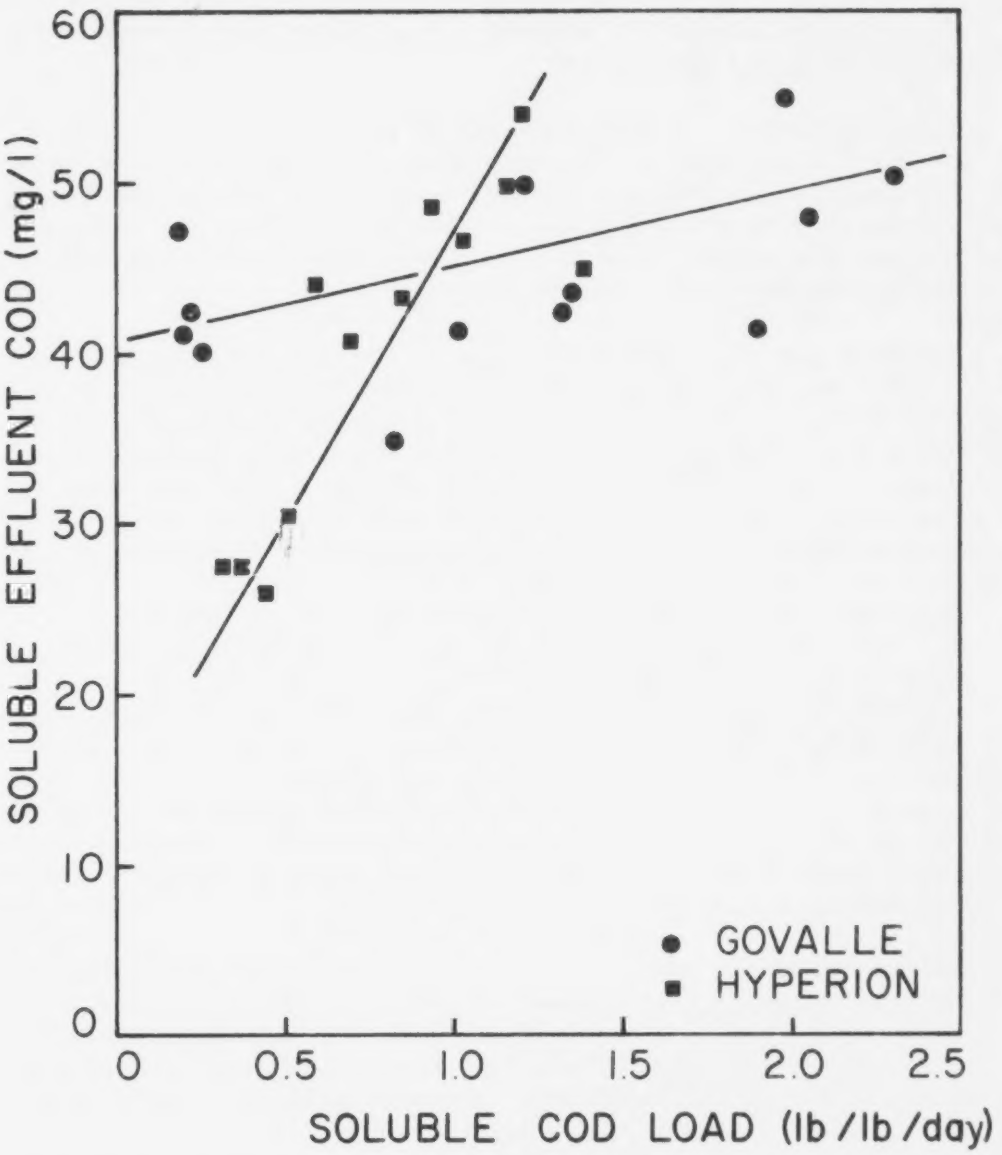


FIG. 3-6
SOLUBLE COD LOAD AND EFFLUENT

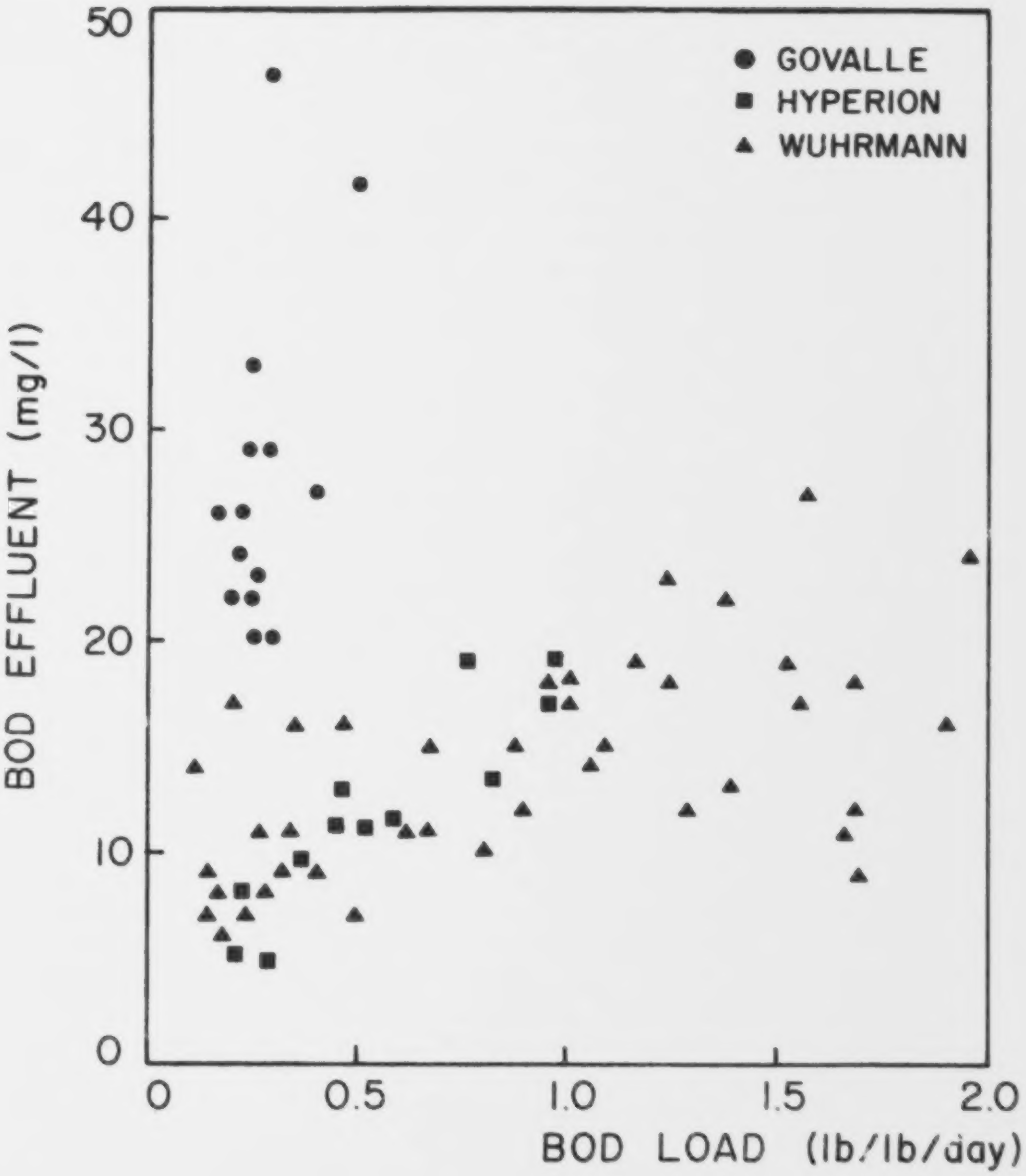


FIG.3-7
BOD LOAD AND BOD EFFLUENT

and 125 minutes, respectively. The mixed liquor suspended solids concentrations were maintained at 3300 and 6000 mg/l, respectively. The overflow rates to the clarifiers were about 530 gallons/sq ft - day which was about the same as that reported for the Hyperion experiments. Therefore, the data reported by Wuhrmann compare favorably with that reported by Hyperion.

The data presented in Figure 3-5 indicated that the soluble effluent BOD concentrations only increased slightly with increasing loading. Therefore, it can be expected that the higher total effluent BOD concentrations are caused primarily by the increased suspended solids concentrations in the effluent. The effluent suspended solids are plotted versus the effluent BOD in Figure 3-8. These data indicate that the effluent BOD concentration increases as the suspended solids concentration increases. The scatter in this particular plot is caused by the number of factors including nitrification in the BOD bottle which results in a relatively high BOD at relatively low suspended solids concentration and the activity of the suspended solids. The suspended solids of a highly loaded plant will exert a higher BOD in the effluent than those solids from a relatively lightly loaded plant. For these data the ratio of the concentrations of total BOD to the suspended solids (BOD/SS) is in the range of 0.55 to 1.10.

The concentration of suspended solids in the clarifier effluent is affected by the hydraulic overflow rate, the solids loading rate to the clarifier, and the settling and flocculation characteristics of the sludge. At a constant overflow rate and initial suspended solids concentration, the settling and flocculation of the sludge are the only factors which influence the effluent concentration. These sludge properties can be related to sludge age. Therefore, the effluent concentration of suspended solids were plotted against sludge age for the data reported by Wuhrmann and for the experiments at the Hyperion Plant since the overflow rate for these experiments was the same (530 gpd/sq ft). These data are presented in Figure 3-9 and indicate that the effluent suspended solids concentration increases as the sludge age decreases. However, the effluent suspended solids concentration would increase sharply for a longer sludge age typical of the extended aeration systems. Temperature is also a factor as indicated by higher effluent suspended solids concentrations at lower temperatures as reported by Wuhrmann.

The performance of the activated sludge process treating municipal wastewater is controlled in some respect by the concentration of suspended solids in the effluent. The suspended solids in the effluent of a well operated secondary clarifier are essentially small floc particles (pin flocs) and single microbial cells. As a concentration of ciliated protozoa increases, the effluent suspended solids decrease. These protozoa are predators of bacteria and consume the fine floc particles. The growth rate of the protozoa in most cases is much lower than that for the bacteria. Therefore, as the sludge age increases, there is a better chance for the protozoa concentration to

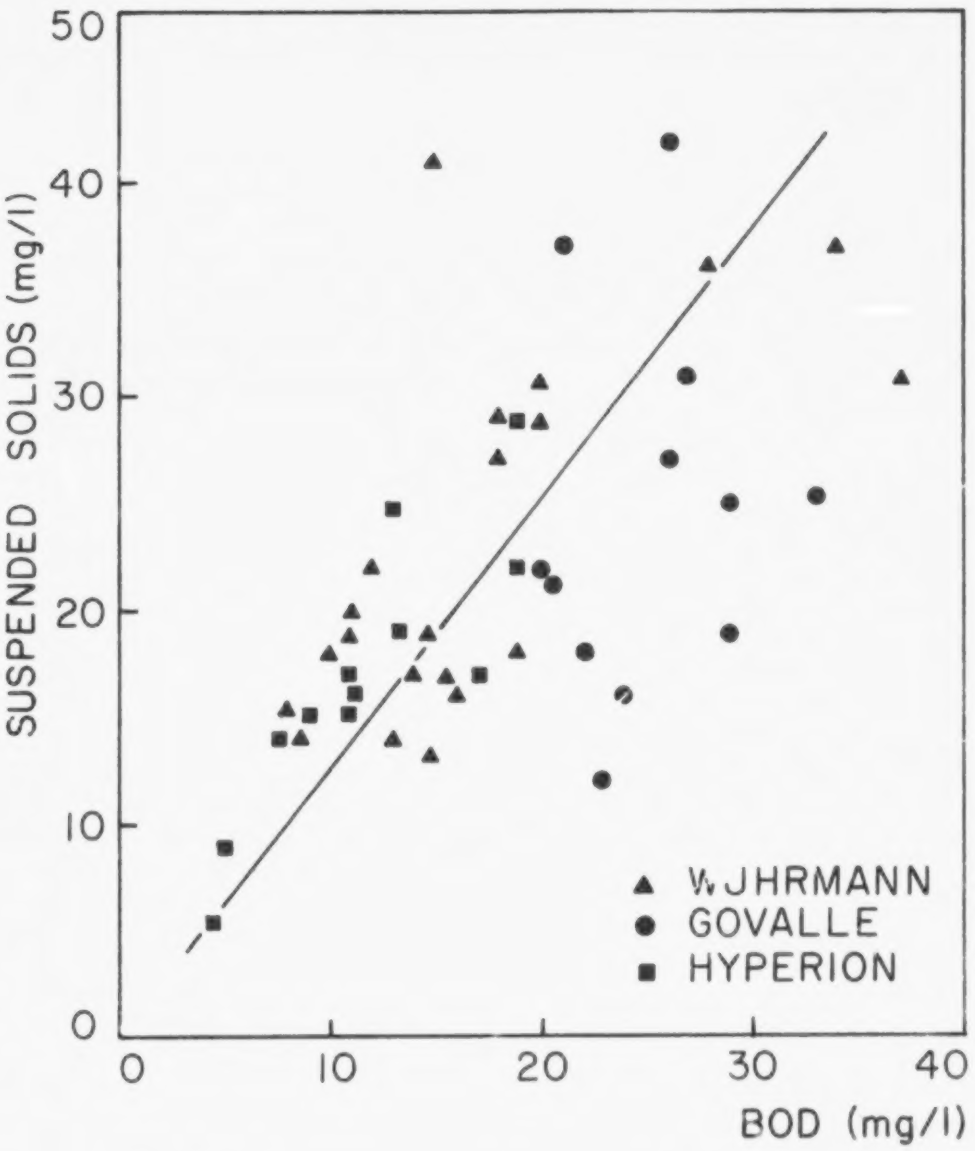


FIG. 3-8

BOD AND SUSPENDED SOLIDS EFFLUENT

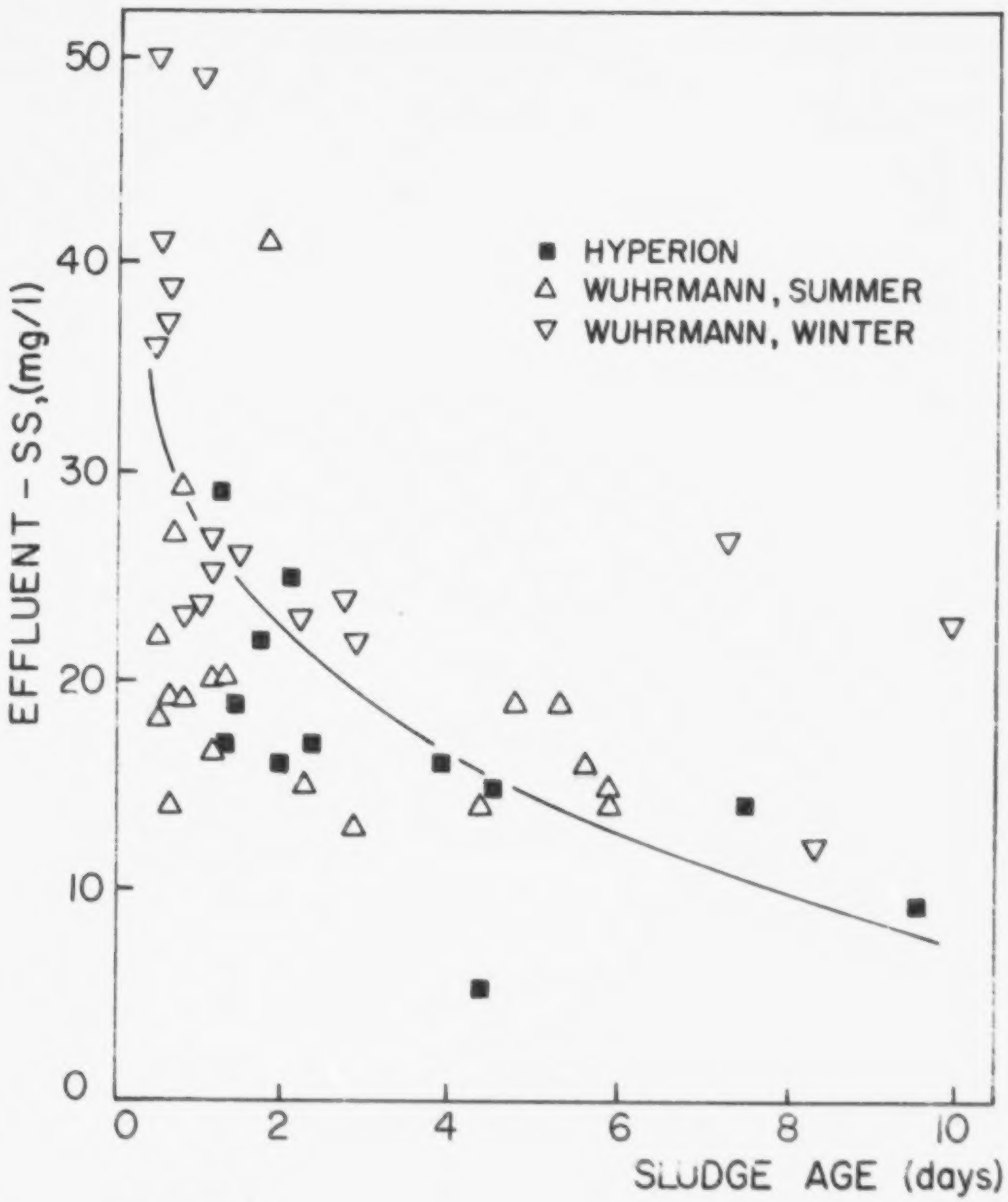


FIG.3-9
EFFLUENT SS AND SLUDGE AGE

increase resulting in a reduction in the concentration of suspended solids in the effluent and an improved effluent quality. Temperature also affects the performance of a clarifier. As the temperature decreases, the viscosity of the water increases and thereby reduces the settling rate of the floc particles. Therefore, a higher concentration of suspended solids in the effluent is observed at low temperatures.

Lower overflow rates in the final clarifier did not seem to be the only solution to increasing the efficiency of the final settling process. Other remedial procedures may be dictated by the characteristics of the sludge. In some cases, flocculating agents may be required to produce a more stable and more settleable floc. The gases resulting from denitrification or from the capture of air bubbles may cause floating floc particles, and some way of releasing the gas is required to improve settling. Final clarifiers may be plagued with density currents which affect the efficiency of solids removal by possibly carrying suspended solids from the sludge blanket up over the effluent weirs.

The mathematical model presented in Equation 3-3 may be used to calculate oxygen requirements if no nitrification takes place.

$$\frac{R}{X} = a' \left(\frac{S_o - S_e^*}{Xt}\right) + b' \qquad 3\text{-}3$$

In practice the oxygen requirements per unit volume may be more useful. This equation may be modified to include the oxygen requirements for nitrification.

$$R = \frac{8.34}{1000} \quad a' \frac{24\,(S_o - S_e^*)}{t} + b'X + 4.60 \frac{24 \Delta NH_3}{t} \qquad 3\text{-}4$$

in which:

R = oxygen uptake per unit volume (lb/1000 gal/day)

t = theoretical aeration time (hours)

S_o = influent BOD (mg/l)

S_e^* = soluble effluent BOD (mg/l)

X = MLSS (mg/l)

ΔNH_3 = ammonia removal (mg/l)

a' = constant relating to the oxygen required for sludge synthesis (for municipal wastewater = 0.53 lb/lb)

b' = endogenous oxygen uptake rate (for municipal wastewater = 0.15 lb/lb/day).

In most municipal biological treatment plants the residual soluble effluent BOD is three to seven mg/l which in fact is about five percent of the influent concentration, and $S_o - S_e^* = 0.95S_o$. Therefore, a' = 0.53 lb/lb, and $a'(S_o - S_e^*) = 0.05S_o$. Therefore, equation 3-4 may be rewritten as Equation 3-5.

$$R = \frac{8.34}{1000}\left[0.5\frac{24S_o}{t} + 0.15X + 4.6\frac{24\Delta NH_3}{t}\right] \qquad 3\text{-}5$$

These equations can be used to calculate the total oxygen requirements per unit volume of aeration tank per day. However, the distribution of oxygen required to satisfy the oxygen uptake rates which vary throughout the aeration tank and throughout the course of the day is not provided. The maximum oxygen uptake rate was observed in laboratory and full-scale experiments to occur at the end of the daily peak loading. Therefore, the oxygen uptake rate reaches a peak normally in late afternoon when the rate of flow and the strength of the incoming wastewater begin to decrease. The ratio of the peak oxygen uptake rate to the daily average oxygen uptake rate is much higher in those plants which are highly loaded than in the plants which have a low organic loading. In the low loaded plants, the endogenous oxygen uptake rate tends to equalize the peak oxygen uptake rates which may develop. The ratio of oxygen uptake rates is also lower than the ratio of the peak organic loading to the average organic loading for a given day. The results of laboratory-scale studies indicate that some of the suspended solids which accumulate in the mixed liquor undergo degradation after the peak organic loading has passed through the plant (Berryhill, 1970). This oxygen requirement caused by the degradation of the insoluble organic material may be attributable to the fact that these organic solids undergo degradation at a much lower rate than the dissolved organic material. Therefore, for design purposes, it can be assumed that the distribution of oxygen uptake rates at a treatment plant is similar to the distribution of incoming wastewater flow. The oxygen uptake rate also varies with the length of longitudinal tanks. The biodegradation of soluble organic material is high compared to that for the nonsoluble material. Therefore, the oxygen uptake rate in the aeration tank is almost always higher in the first part of the tank. The oxygen uptake rate will then decrease as the concentration of soluble organic material decreases. The results of experimental studies point to this variation in oxygen uptake rates and the peaks that can occur. These data presented in Figure 3-10 indicate the relationship of oxygen uptake rate to aeration time for a batch contact

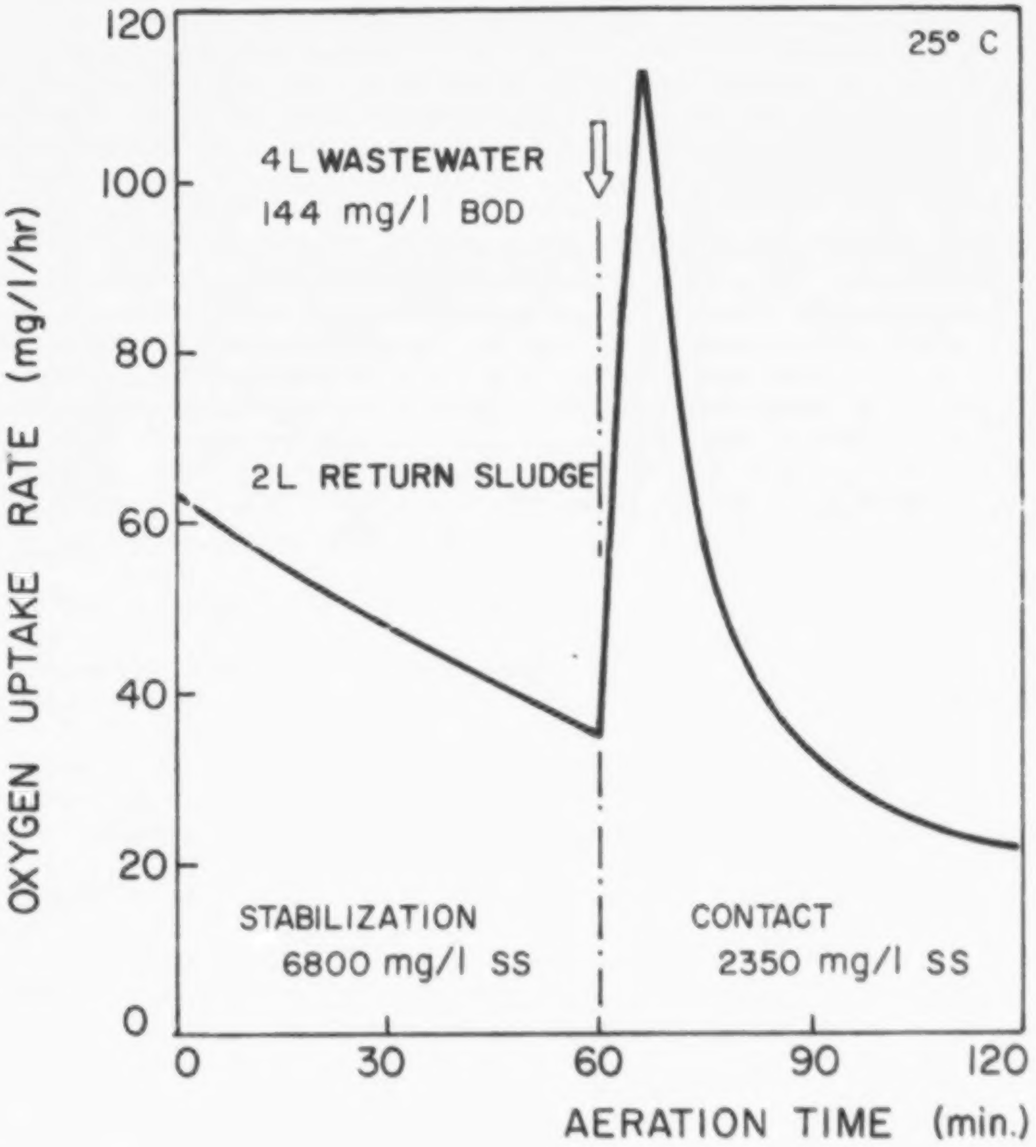


FIG.3-10
OXYGEN UPTAKE RATES (BATCH)

stabilization system. After the wastewater is introduced, the oxygen uptake rate increases sharply until a peak is reached and decreases markedly with time. This peak coincides with the rate of utilization of the soluble organic material. In the full-scale plant where the contact and stabilization zones were not physically separated, this peaking of oxygen uptake rate was reduced since some degree of backmixing was observed. This backmixing causes mixing of the influent with a larger volume of mixed liquor and tends to equalize the oxygen uptake rate.

The distribution of incoming wastewater to various points in the aeration tank in the step aeration process also tends to maintain a relatively constant oxygen uptake throughout the entire length of the tank. The results of oxygen uptake profiles determined in the contact stabilization plant and the conventional activated sludge system at the Govalle Treatment Plant in Austin, Texas are shown in Figure 3-11. These results indicate a similar pattern, namely that the oxygen uptake rate is highest shortly after the influent wastewater is mixed with return sludge and decreases rapidly to an equilibrium level. After this point, the oxygen uptake rate decreases slightly to a minimum rate at the effluent of the aeration basin.

AERATION

The oxygen required for biodegradation of organic material in wastewaters is usually introduced into the aeration tank as diffused air or by mechanical equipment. The aeration equipment is normally evaluated under standard conditions of a water temperature of $20^{\circ}C$ and an oxygen concentration of zero. The aeration equipment is classified according to oxygenation capacity (OC) expressed in units of pounds of oxygen per unit volume of aeration tank (lb O_2/1000 gal-day). At times, the aeration efficiency (N_o) is also provided in terms of pounds of oxygen per horsepower hour (lb/HP-hr). However, the oxygen transfer is different under process conditions than under standard conditions. An oxygen concentration of about two mg/l is generally maintained in the aeration basin although it may be possible to operate at lower dissolved oxygen concentrations.

The oxygen transfer rate constant ($k_L a$) is related to the oxygenation capacity by Equation 3-6.

$$OC = (k_L a)(C_s) \qquad 3\text{-}6$$

in which:

C_s = concentration of oxygen in water at saturation

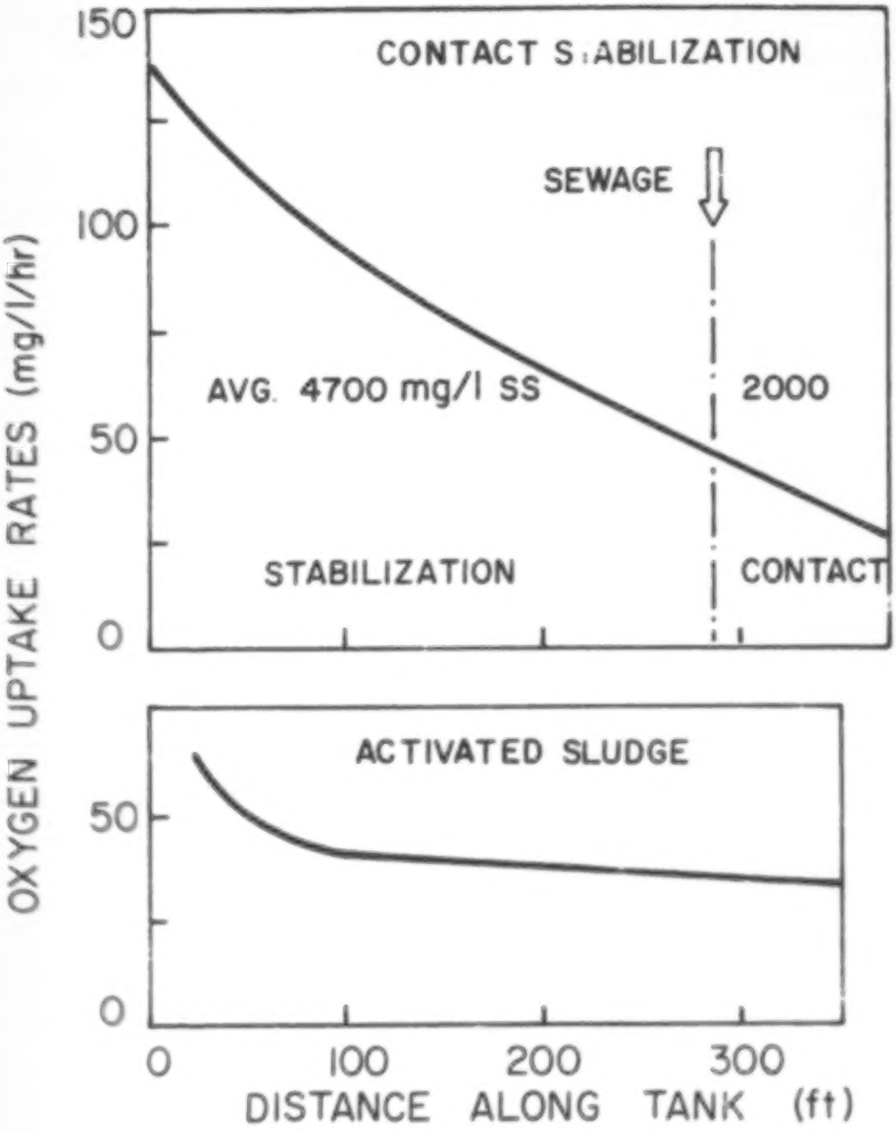


FIG.3-11
OXYGEN UPTAKE PROFILES

The oxygen transfer rate constant increases as the temperature increases, but the oxygen concentration at saturation decreases at higher temperatures. Therefore, although these two parameters are affected by temperature, the oxygenation capacity is relatively independent of temperature. The oxygenation capacity decreases by about six percent when the temperature increases from five degrees to 30°C.

The oxygen transfer is different for wastewater, for mixed liquor and for tap water under comparable conditions. The actual oxygen transfer rate, therefore, can be expressed by Equation 3-7:

$$OT_T = OC \frac{C_{s_T} - C}{C_{s_{20}}} \eta \, 1.02^{(20-T)} \qquad 3\text{-}7$$

in which:

OT_T = oxygen transfer rate at temperature T in °C (lb/1000 gal-day)

OC = oxygenation capacity at T = 20°C (lb/1000 gal-day)

C = concentration of oxygen in liquid (mg/l)

C_{s_T} = concentration of oxygen in liquid at saturation at T°C (mg/l)

$C_{s_{20}}$ = concentration of oxygen in liquid at saturation at T=20°C

η = ratio of oxygen transfer under process conditions to the oxygen transfer in tap water at the same temperature

The effect of temperature on the oxygenation capacity is presented in Figure 3-12. The curve shows that temperature changes have a more noticeable effect at the lower temperatures than at the higher temperatures.

The concentration of oxygen at saturation in a diffused air aeration system must be corrected for the increased pressure that occurs as the air is released at the bottom of the tank. The saturation concentration of dissolved oxygen can be adjusted for the partial pressure at the mid-depth of the tank by using Equation 3-8.

$$C_{sm} = C_s \left(\frac{P_b}{29.4} + \frac{O_t}{42}\right) \qquad 3\text{-}8$$

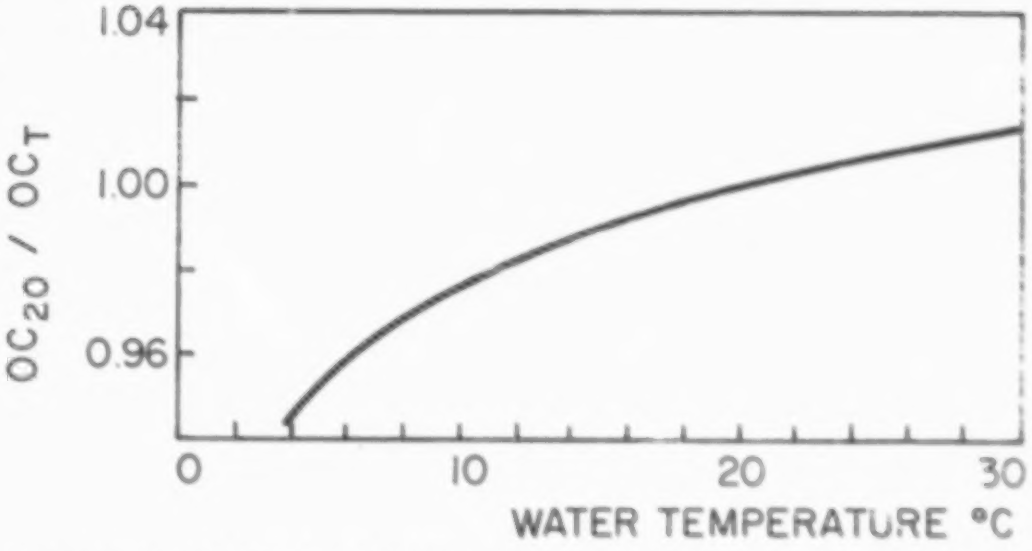


FIG.3-12 INFLUENCE OF TEMPERATURE ON OXYGENATION CAPACITY

in which:

C_{sm} = concentration of oxygen at saturation at mid-depth (mg/l)

C_s = concentration of oxygen at saturation under standard conditions (mg/l)

P_b = absolute pressure at depth of air release (lb/sq in)

O_t = oxygen in air leaving the aeration tank (percent)

The concentration of oxygen at saturation at mid-depth (C_{sm}) therefore depends on the depth of the diffuser as well as the percentage of oxygen in the air leaving the aeration tanks. The effect of the depth at which the diffusers are located on the ratio of the concentration of oxygen at saturation at mid-depth to the concentration of dissolved oxygen at saturations under standard conditions is presented in Figure 3-13. This curve was calculated for an oxygen transfer efficiency of five percent and indicates a linear relationship between the ratio C_{sm}/C and depth.

Various values of alpha used in Equation 3-7 have been reported for municipal wastewater. In bubble aeration systems, alpha has been reported to be equal to between 0.60 and 0.80 whereas in surface aeration systems, this value ranged from 0.80 to 1.10.

The interaction between oxygen transfer, depth and rate of air flow are presented schematically in Figure 3-14. A linear relationship between the oxygen transfer and the rate of air flow at a constant depth exists. However, at the extremely high air flow rates, the rate of oxygen transfer tends to level off at some equilibrium value. There is also a linear relationship between oxygen transfer and depth at a constant air flow rate. The exact value of the oxygen transfer at different depths and air flow rates is controlled to a large extent by the type of diffuser or bubble introduction device used. The aerator configuration and tank geometry influence the oxygen transfer rate. In a spiral flow tank, the actual velocity of the water increases with increasing air flow rates; therefore, the water air bubble contact time and the shear of the bubbles decrease resulting in a lower oxygen transfer per unit of volume of air. The ratio of the depth to width of the aeration tank also influences the oxygen transfer rate. The aeration device controls the diameter of the bubbles which are released. The oxygen transfer decreases as the diameter of the bubble increases. The oxygen transfer for a specific aeration device can be expressed as the oxygen transfer per 1000

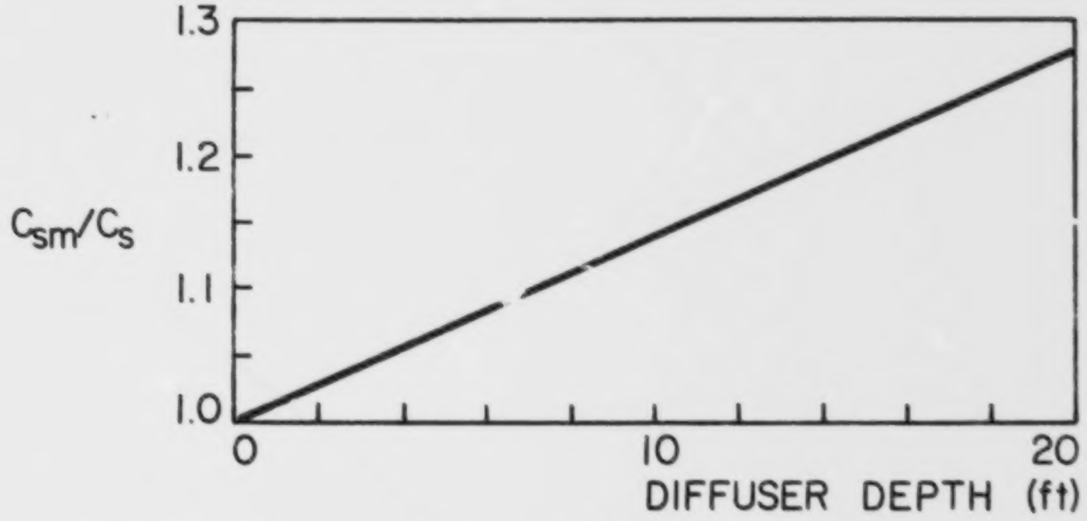


FIG.3-13 OXYGEN SATURATION AND DEPTH OF AIR RELEASE

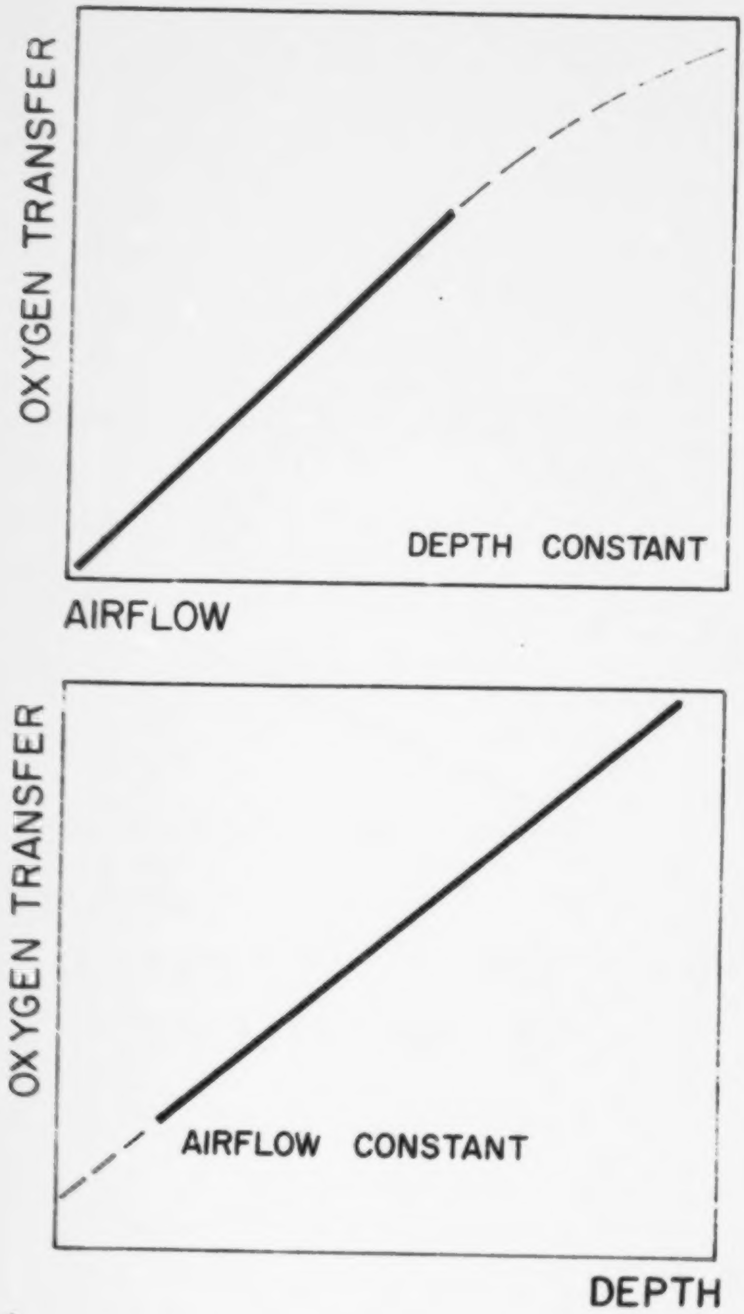


FIG.3-14
BUBBLE AERATION — OXYGEN TRANSFER
(SCHEMATIC)

cubic feet of air flow per foot of depth of the air release. Each aeration device has optimum air flow rate beyond which the aeration efficiency decreases.

Mechanical aerators which are located at or near the surface of the liquid in the aeration tank are being used to a large extent. These aerators can be categorized as:

(a) cone aerators with or without draft tubes

(b) pump-type aerators which operate at high speeds, and

(c) brush aerators.

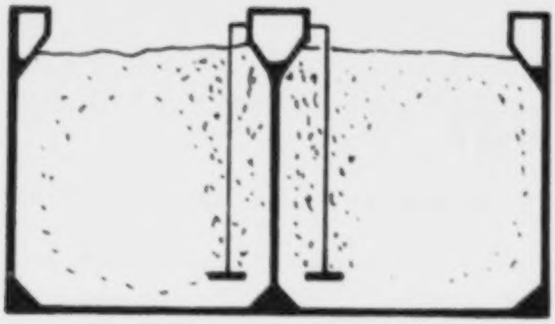
Schematic diagrams of the various types of surface aerators are shown in Figure 3-15.

Pump-type aerators are directly coupled to the motor and no gear box for speed reduction is required. The oxygen transfer can be controlled by introducing air into the suction portion of the aerators. However, less water is pumped.

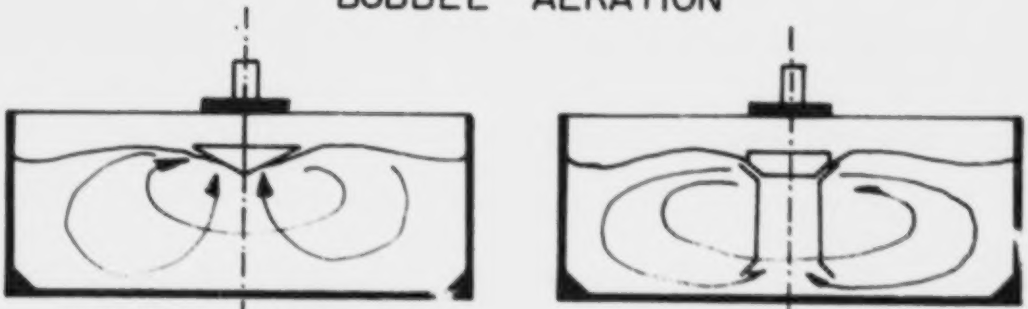
The oxygen transfer of cone aerators increases within a given range with the depth of immersion of the aerator and with increasing peripheral velocity. Available data indicate that the increase in oxygen transfer is proportional to the third power of the peripheral velocity. The oxygen transfer also increases as the tank volume is reduced. Therefore, the oxygenation efficiency expressed as lb/HP-hr increases with increasing power level expressed as HP/1000 gallons. The general relationships between oxygen transfer and depth of immersion and the peripheral speed are illustrated in Figure 3-16. The relationship between oxygen transfer efficiency and power level is also illustrated schematically. The characteristics of aeration device determine the need for a draft tube. Draft tube aerators usually can be used only in very shallow tanks.

Sludge deposits may develop in the aeration tank when mechanical aerators are used. Knop and Kalbskopf (1968) indicate that the velocity at the bottom of an aeration tank in which cone or turbine-type aerators were used is a function of a power level. It has been observed that sludge deposits could not be found in tanks where the bottom velocity was much lower than one ft/sec. These reports indicate that turbulent conditions exist in the bottom of the tanks and that the velocities cannot be effectively measured by the common propeller-type meters.

Brush aerators which were developed by Kessener were initially mounted on the walls of aeration tanks in which spiral flow was maintained. However,

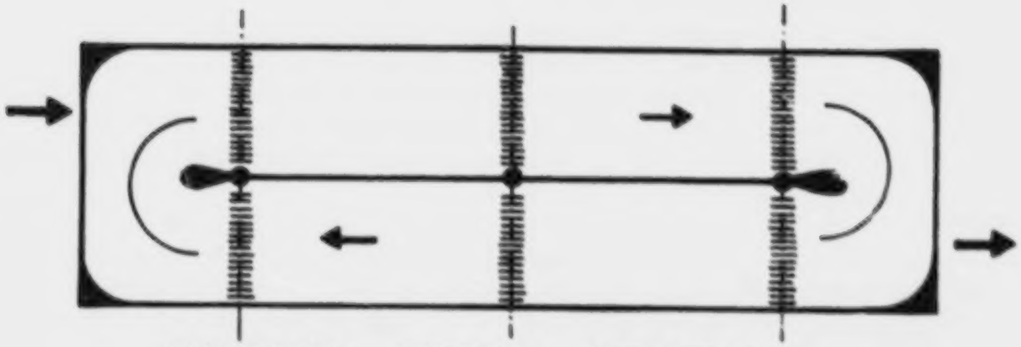


BUBBLE AERATION



WITHOUT DRAFT TUBE WITH DRAFT TUBE

CONE AERATION



MAMMOTH ROTOR AERATION

FIG.3-15

TYPICAL AERATION TANKS

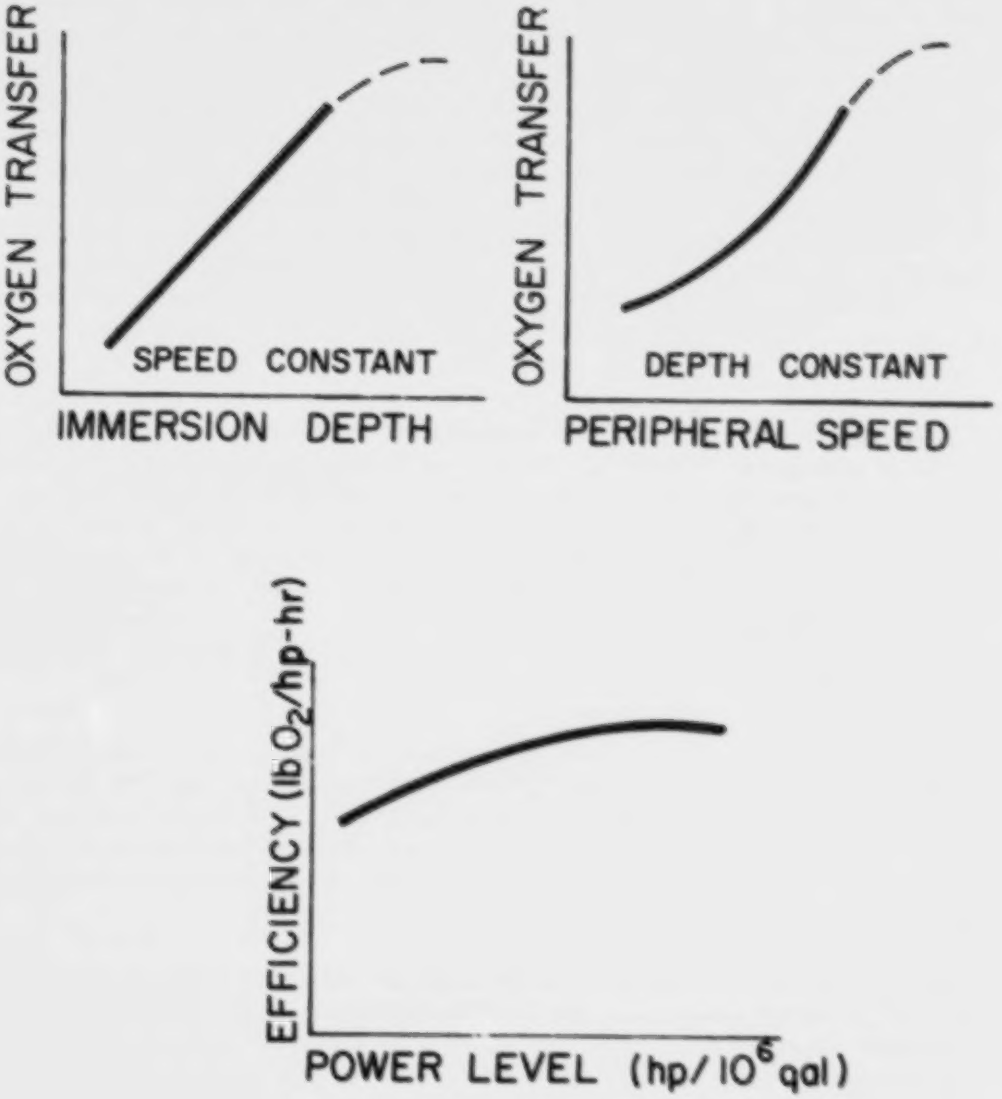


FIG.3-16

MECHANICAL AERATOR CHARACTERISTICS

the development of the cage-rotor and the mammoth rotor has eliminated the use of wall mounted brushes. Brushes are now mounted in a horizontal position perpendicular to the direction of flow. If more than one pair of rotors is operated in a single tank, baffles must be installed near the rotors to enhance mixing and maintain higher oxygen transfer rates. The oxygen transfer for brush aerators increases linearly with the depth of immersion and with the peripheral speed raised to the 2.5 power. The brushes are usually operated at a fixed speed and the oxygen transfer can be controlled by the depth of immersion. The rate of oxygen transfer to an aeration system of a specific surface area can be reported as pounds of oxygen per hour per unit or pounds of oxygen per HP/hour. The operating range must be given and the dimensions of the tank must be provided in order to effectively design a system using brush aerators.

Von der Emde and Kayser (1969) reported the oxygenation capacity and oxygen transfer efficiency of various aeration devices under process conditions. These data indicate that for bubble aeration systems in which nozzles with 0.28-inch diameter orifices were used, the value for OC = 0.95 lb O_2/1000 cu ft air/ft of depth of air release and a value of N = 1.30 lb O_2/HP-hr. The data reported for ceramic tube-type diffused aeration systems was OC = 0..2 lb O_2/1000 cu ft air/ft of depth of air release and N = 1.85 lb O_2/HP-hr. The oxygen transfer efficiency for surface aerators of the cone-type was reported as N = 2.0 to 3.5 lb O_2/HP-hr.

The above discussion of aeration equipment indicates that each aeration device requires a specific and different tank geometry, and the selection of aeration equipment must include the cost of the tanks as well as the aeration efficiency under process conditions. More detailed discussion of the design of aeration equipment may be found in Gloyna and Eckenfelder (1968) and in Eckenfelder and Ford (1970).

Recently the use of pure oxygen for aeration has been investigated by Albertson (1970). Two full-scale activated sludge treatment plants were operated in parallel in which one system was aerated with pure oxygen and the other with diffused air. However, some of the operating conditions such as sludge age and mixed liquor volatile suspended solids concentration were different in the air and oxygen systems operating in parallel. The oxygen system operated at a higher sludge age and mixed liquor volatile suspended solids concentration than the air system. The increased sludge age results in a lower production of cells, a greater destruction of volatile solids and an overall lower sludge yield. The quantity of soluble BOD removed per unit tank volume increases as the concentration of mixed liquor volatile suspended solids increases. These factors partially explain the higher rate of BOD removal and a lower sludge yield reported for the oxygen system when compared with the aerated activated sludge system. Ball and

Humenick (1971) prepared a review and analysis of oxygen systems for municipal wastewater treatment and concluded that although oxygen systems may provide some real advantages over air systems, the benefits may not outweigh the costs.

The greatest savings for the pure oxygen system in total annual cost were based on the low production of solids and the reduced sludge handling costs. A comparison of the pure oxygen system with surface aerators which in fact have on the average about a 50 percent higher efficiency of oxygen transfer than the diffused air system may have resulted in different results.

WASTE SLUDGE AND RETURN SLUDGE

The quantity of sludge produced during the activated sludge process is a function of the sludge age as well as the quantity of BOD removed and can be expressed mathematically as:

$$\frac{1}{G} = a^*\left(\frac{S_o - S^*_e}{xt}\right) - b + \frac{X_{on}}{Xt} \qquad 3\text{-}9$$

Typical values of a* and b are 0.63 and 0.075, respectively. The soluble BOD in the effluent is equal to about five percent of the initial BOD or $S^*_e = 0.05\ S_o$. The results of two studies using municipal wastewater indicate that the nonbiodegradable influent suspended solids represented 60 percent of the influent suspended solids. Introducing these values into Equation 3-9 results in Equation 3-10 which can be further reduced to Equation 3-11.

$$\frac{1}{G} = 0.63\left(\frac{0.95\ S_o}{Xt}\right) - 0.075 + \frac{0.6\ X_o}{Xt} \qquad 3\text{-}10$$

$$\frac{1}{G} = \frac{0.6\ (S_o + X_o)}{Xt} - 0.075 \qquad 3\text{-}11$$

The relationship presented in Equation 3-11 between the sludge age and the summation of the influent BOD and suspended solids is presented in Figure 3-17. The data points are those reported by Wuhrmann (1964) for pilot plant studies as well as for the Hyperion and Govalle full-scale experiments. The line drawn through the points has a slope of 0.60 and provides a basis for estimating the amount of sludge produced in a somewhat better fashion than the growth models which depend on the composition of the wastewater and for which different values of the coefficients a and b must be determined. The scatter in the data, especially in the results reported for the operations at the Govalle Treatment Plant may be explained partially by the fact that the quantity of sludge wasted at this plant is merely an estimate and is not based on any specific flow measurements. There is also a greater percentage of nonbiodegradable solids in the effluent to the Govalle Plants, since no primary clarification preceeds the aeration tank. However, the data reported by Hyperion are described quite well by the line shown on Figure 3-17. This closeness of fit may be explained by the fact that at the Hyperion Plant, primary treatment preceeds the aeration tank and much of the nonbiodegradable suspended solids are removed in the primary clarifier. The scatter of the Wuhrmann data can be attributed to difficulties encountered in attempting to measure waste sludge flows in pilot plant operations.

The excess sludge calculated as the reciprocal of sludge age is the total mass of solids leaving the activated sludge system. Therefore, the flow of waste sludge may be calculated.

$$G = \frac{VX}{WX_{RS} + QX_e} \qquad 3\text{-}12$$

$$W = \frac{1}{X_{RS}} \left(\frac{VX}{G} - QX_e \right) \qquad 3\text{-}13$$

The sludge wasting facility should be able to discharge the total amount of sludge wasted per day during a 10 - 14 hour period if we assume that the effluent suspended solids concentration is zero and that the concentration of return sludge is the lowest which can be obtained with the total return sludge pump capacity and a mixed liquor suspended solids concentration of 2,000 mg/l. The return sludge concentration depends on the flow rates of return sludge and of incoming wastewater as well as the concentration of mixed liquor suspended solids maintained in the aeration tank. This relationship can be expressed in Equation 3-14 and 3-15.

$$X_{RS}Q_{RS} = X(Q + Q_{RS}) \qquad 3\text{-}14$$

$$X = X_{RS} \frac{Q_{RS}}{(Q + Q_{RS})} \qquad 3\text{-}15$$

in which:

X = mixed liquor suspended solids concentration (mg/l)

X_{RS} = concentration of suspended solids in the return sludge (mg/l)

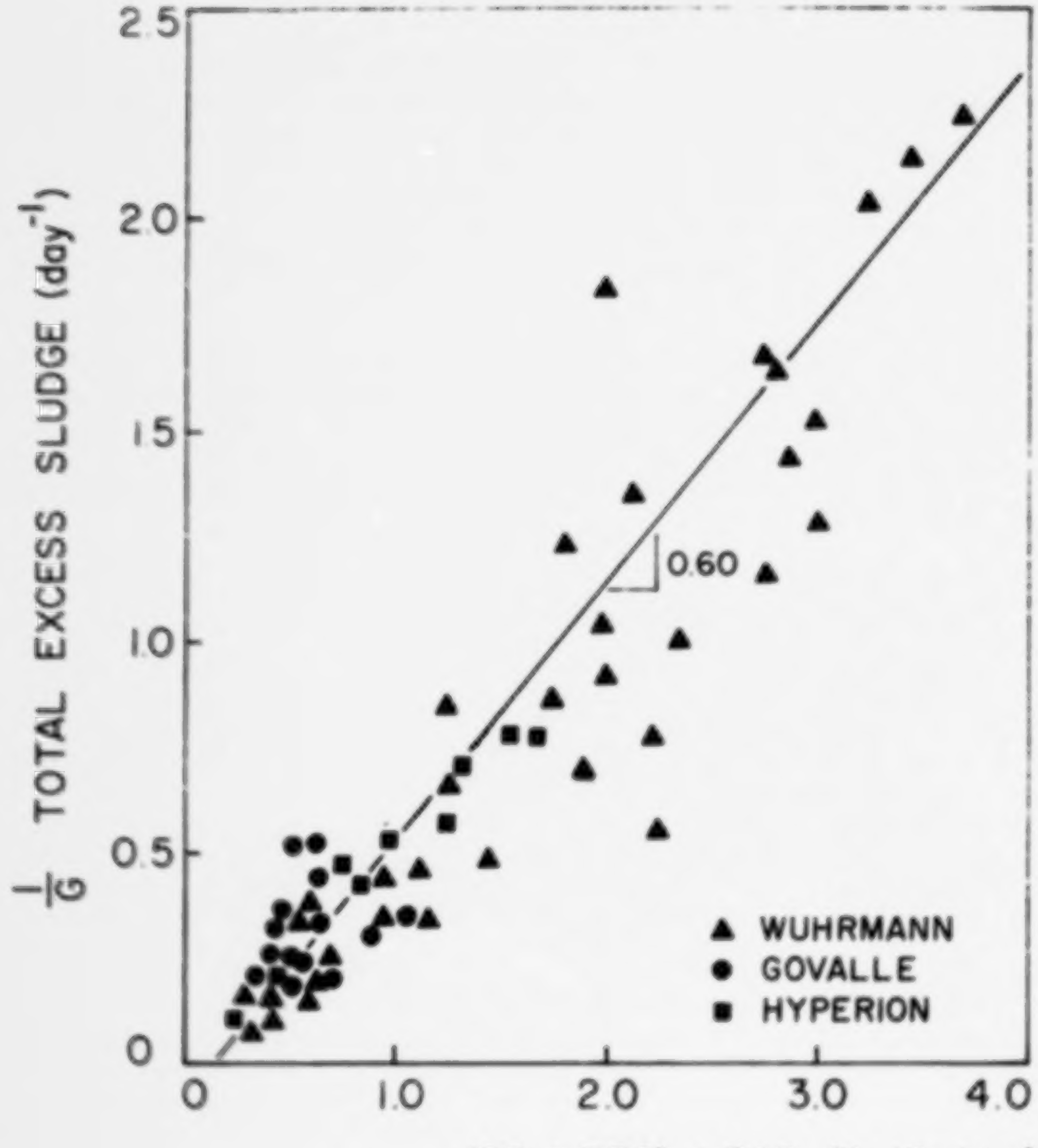


FIG. 3-17

EXCESS SLUDGE AND (SS + BOD) - LOAD

Q_{RS} = return sludge flow rate (1000 gal/hr)

Q = wastewater flow rate (1000 gal/hr)

The concentration of suspended solids in the mixed liquor in the aeration tank will decrease as the incoming wastewater flow increases, if the rate of flow of return sludge and the concentration of suspended solids in the return sludge remain constant. However, the variations in the concentration of mixed liquor suspended solids will decrease as the ratio of return sludge flow to incoming wastewater flow increases. It is from this point of view that the return sludge flow rate should be maintained at some reasonable level or not be permitted to be very low.

The concentration of suspended solids in the return sludge is generally estimated by using the concentration of activated sludge which results after 30 minutes of settling. The reciprocal of this concentration of solids in the settled sludge is the sludge volume index. Therefore, the concentration of suspended solids in the return sludge may be expressed as Equation 3-16.

$$X_{RS} = \frac{10^6}{SVI} \qquad 3\text{-}16$$

in which:

X_{RS} = concentration of suspended solids in the return sludge in mg/l

SVI = sludge volume index (milliliters per gram) and represents the volume occupied by one gram of sludge after settling for 30 minutes in a one liter cylinder.

The sludge volume index for a municipal activated sludge plant is about 150. The limiting concentration of suspended solids in the return sludge will therefore be about 6,700 mg/l. At a return sludge flow equal to the incoming wastewater flow, a mixed liquor suspended solids concentration of about 3500 mg/l could be maintained. Therefore, in general the capacity of the return sludge pumps should be about equal to the average wastewater design flow. The aeration tank should be designed based on a concentration of suspended solids in the mixed liquor of about 3,000 mg/l.

The energy required for pumping the return sludge is negligible compared to the energy required for aeration. Therefore, the return sludge pumps could be operated at a constant flow rate of the entire 24-hour period. By operating continuously at a predetermined flow rate, the return sludge system would not require any elaborate flow control devices. Two or three return sludge pumps having the same capacity should provide enough flexibility to handle the variations of return sludge flow required. In most municipal treatment plants, it is common practice to have more than one return sludge pump. Screw pumps are preferred in the newer plants in Europe. In many cases, a single screw pump is installed and is capable of returning the quantity of sludge required at the large treatment plants.

It is difficult to maintain and control the concentration of suspended solids in the mixed liquor by changing the rate of return sludge flow. A more effective way of controlling the concentration of mixed liquor suspended solids is by wasting sludge. Unfortunately, this method provides for only a decrease in the concentration of the mixed liquor suspended solids. Therefore, wasting sludge should be carefully controlled. An alternate system is to waste sludge continuously; however, if semi-continuous wastage is required, this waste can be controlled by pumps which are operated on a time cycle or by using some other suitable flow control device.

EXTENDED AERATION PROCESS

The extended aeration process is a modification of the activated sludge process in which the sludge age is maintained at a relatively high value providing time for a portion of the sludge to be stabilized. An extended aeration plant normally consists of an aeration tank, followed by a final clarifier and the necessary pumps for sludge return. The excess sludge is generally wasted in the form of suspended solids in the effluent.

The extended aeration process is widely used for relatively small communities in the United States and Europe. The basic difference in the operation of extended aeration plants in the United States and Europe is that in Europe the excess sludge is removed from the system in order to maintain effluent which contains a BOD of five to 15 mg/l. At the small plants serving communities of 100 to 500 people, the sludge is dewatered on the sludge drying beds. At the larger plants, the excess sludge is stored and subsequently trucked to the agriculture areas. The basic design parameter is the degree of stabilization of the waste sludge. At organic loadings of about 0.05 lb BOD/lb of MLSS-day, the sludge is reported to be well stabilized. Storage of this material under anaerobic conditions does not cause any severe odor problems. A number of plants are operated at organic loadings of 0.1 lb BOD/lb MLSS-day in order to reduce the volume of aeration that is required. The excess sludge from these plants is partially stabilized. The organic loading rate expressed in terms of initial suspended solids concentration as well as initial BOD concentration can be calculated using Equation 3-11. If one assumes that the wasted sludge is well stabilized at a sludge age of 30 days, this equation can be modified.

$$\frac{1}{\theta} = 0.6\left(\frac{X_o + S_o}{Xt}\right) - 0.075 \qquad 3\text{-}17$$

therefore

$$\frac{X_o + S_o}{Xt} = 0.18 \qquad 3\text{-}18$$

The BOD loading rate would therefore be 0.065 lb BOD/lb MLSS-day based on the assumption that the ratio of influent suspended solids to BOD is about 1.75. The volume of the aeration tank for an extended aeration plant can be based on a load of influent BOD and suspended solids of 0.15. This value would give a somewhat larger aeration volume compared to the value calculated from the theoretical equations.

FINAL CLARIFIERS

The final clarifiers represent an important unit process in the activated sludge plant. The suspended solids are separated from the effluent liquid in the final clarifier and the degree to which the sludge concentrate in the final clarifier will control to some extent the rate at which return sludge must be pumped to the aeration tank.

Clarifiers can be divided into two characteristic groups depending on the hydraulics, namely upflow clarifiers and horizontal clarifiers. The clarifiers may be circular or rectangular and exhibit either flow pattern. Deep clarifiers with a relatively small surface area are generally the upflow types whereas the shallow basins with large surface areas are more likely to be horizontal flow types. The design of final clarifiers is generally based on the overflow rate and the detention time. Theoretical conditions, however, indicate that the overflow rate is far more important than the detention time.

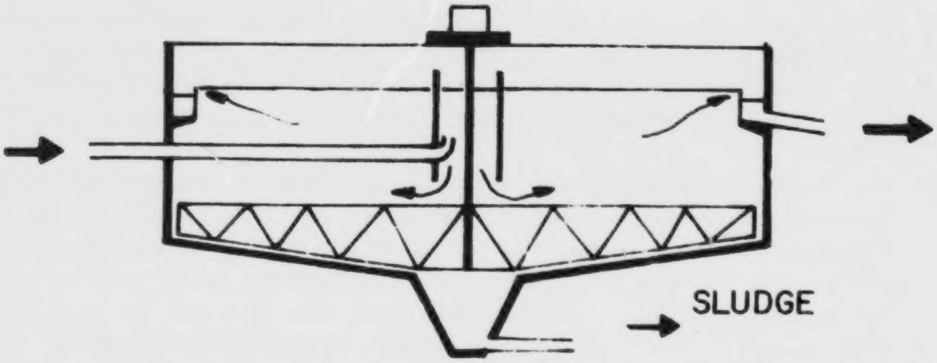
The method by which the settled sludge is withdrawn from the clarifier also affects the performance of the clarifier. The preferred sludge collection mechanism is a vacuum or suction type draw off. The plow type collectors with the chain and flight mechanism in rectangular basins or the bridge with attached plows in circular basins are not very effective in concentrating the waste activated sludge. The sludge tends to flow over the plow as the plow moves through the sludge blanket and very little movement of the sludge to the concentrating compartment takes place. The vacuum or suction type draw off actually draw the solids into the collection pipe under the force of the hydraulic head in the tank. Schematic diagrams of center feed and peripheral feed circular clarifiers and rectangular clarifiers are illustrated in Figure 3-18. The center feed clarifier and the rectangular clarifier illustrate the plow-type collectors whereas the peripheral feed clarifier indicates the suction draw off. Suction draw off equipment can also be installed in the center feed circular clarifier or a rectangular clarifier.

The results of recent investigations by Pflanz (1968) indicate that the effluent suspended solids concentration from final clarifiers operated at constant overflow rates, increased as the concentration of mixed liquor suspended solids increased. The results also indicate that at a constant overflow rate and a constant concentration of mixed liquor suspended solids, the effluent suspended solids concentration increased as the sludge volume index increased. The effluent suspended solids also increased as the temperature decreased.

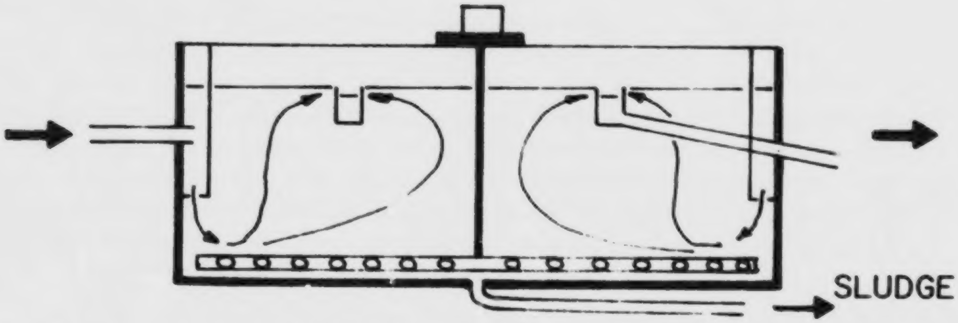
The interdependency of the mixed liquor suspended solids and the sludge volume index indicate that a relationship also exists between effluent suspended solids concentration and the sludge volume surface load. The sludge volume surface load is the product of the overflow rate (gallons per square foot per day) and the fraction of the initial volume occupied by the sludge solids after settling for 30 minutes. Therefore, the sludge volume surface load in a final clarifier with an overflow rate of 1000 gal/sq ft-day, and a sludge which occupies 1/10 the volume of the initial sludge after settling 30 minutes would be 100 gal/sq ft-day.

The relationship between the sludge volume surface load and the effluent suspended solids concentration are presented in Figure 3-19. This plot is based on the data presented by Pflanz (1969). The data show considerable scatter primarily at the higher values of sludge volume surface loading, and effluent suspended solids concentration. However, at the lower range of values, there is a fairly good correlation. The scatter seems to be the result of errors in determining the sludge volume index, which is generally determined under quiescent conditions. However, the conditions in the settling tank are far from quiescent and a good degree of the turbulence in the full-scale tank may actually promote flocculation of the sludge particles. Therefore, in the laboratory determination of the sludge volume index a slow stirring device which is placed in the cylinder would more closely simulate the conditions in the field and may give much more reliable results.

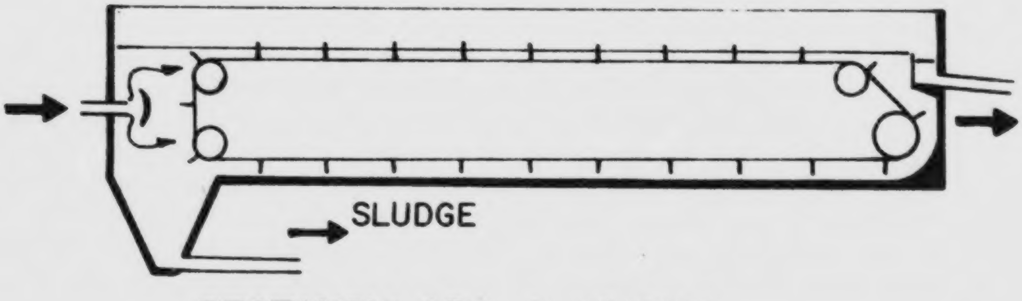
Much of the sludge settles near the inlet zone of the final clarifier under normal overflow flow rates of about 500 gal/sq ft-day and a mixed liquor suspended solids concentration of about 3000 mg/l. A thin layer of diluted sludge may result in a density current. This diluted layer will move along the bottom of the tank, rise to the effluent weirs and result in high effluent suspended solids concentrations. Prevention of these density currents



CIRCULAR CLARIFIER WITH CENTER FEED



CIRCULAR CLARIFIER WITH PERIPHERAL FEED



RECTANGULAR CLARIFIER

FIG. 3-18
DIFFERENT TYPES OF FINAL CLARIFIERS

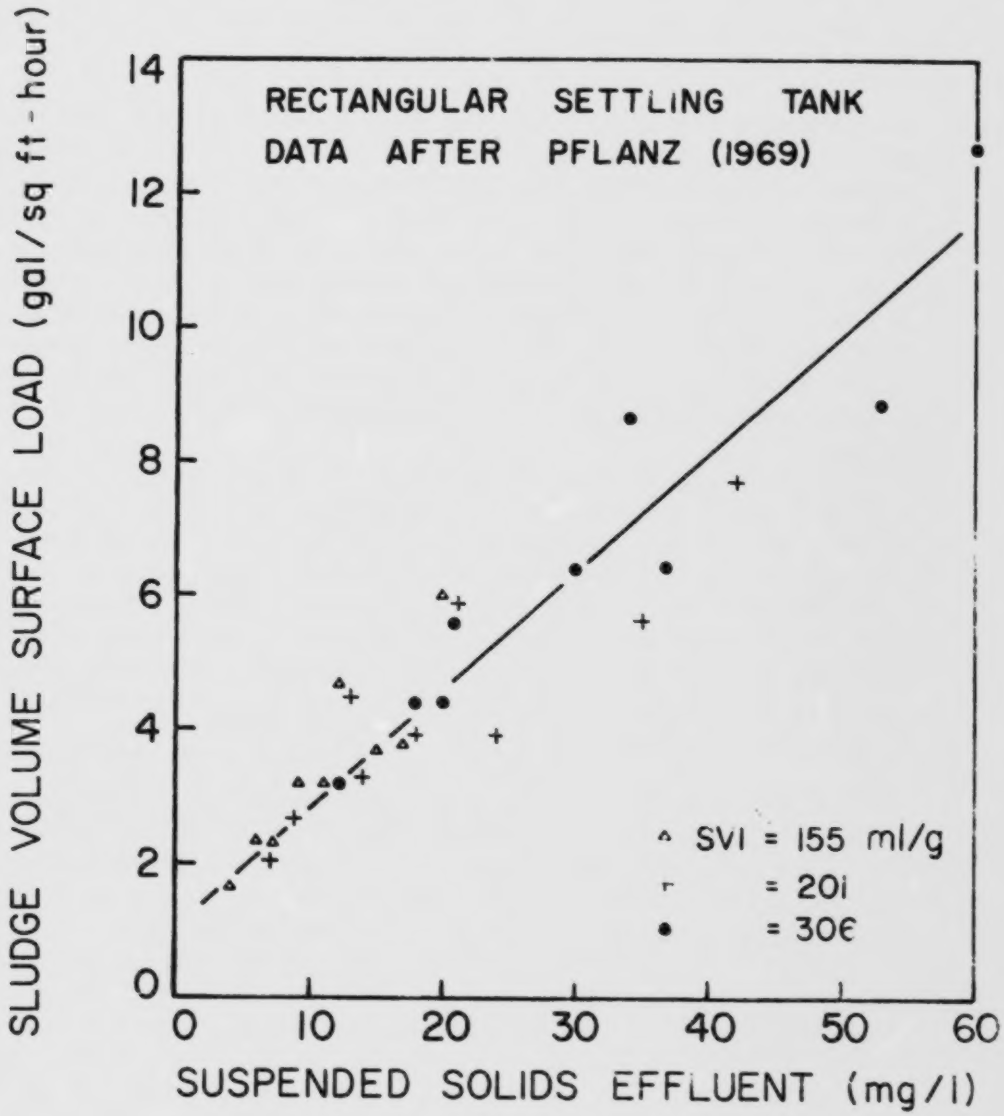


FIG.3-19

SLUDGE VOLUME SURFACE LOAD AND EFFLUENT SUSPENDED SOLIDS would reduce the concentration of suspended solids in the effluent. Density currents were minimized at rectangular final clarifiers which were eight to ten feet deep at three plants in Europe. The method used for minimizing the density currents are illustrated in Figure 3-20.

The settling tanks at the treatment plant at Nordhorn, Germany, had a single effluent weir across the entire width of the tank. However, high effluent suspended solids concentrations were observed and attributable primarily to density currents. A vertical plastic sheet was placed at a point about 1/4 the total tank length and essentially divided the tank into two separate parts. Most of the solids settled in the first part. The increased velocity of flow over the top of the sheet caused the remainder of the suspended solids to become well mixed. Therefore, no density currents were observed in the second part of the tank. This procedure reduced the concentration of suspended solids observed in the effluent.

The results of experiments conducted at the treatment plant at Baden, Austria, indicated that if the weirs at the end of the tank were closed and the effluent was only taken off by the weirs along the side of the tank as illustrated in Figure 3-20, the concentration of suspended solids in the effluent was much less than when all weirs were operated. Similar results were reported at the treatment plant in Ergolz, Switzerland. In this case, six weirs were originally used as those illustrated in Figure 3-20. However, when the weirs 1, 3, and 5 were closed, the concentration of suspended solids in the effluent was markedly reduced.

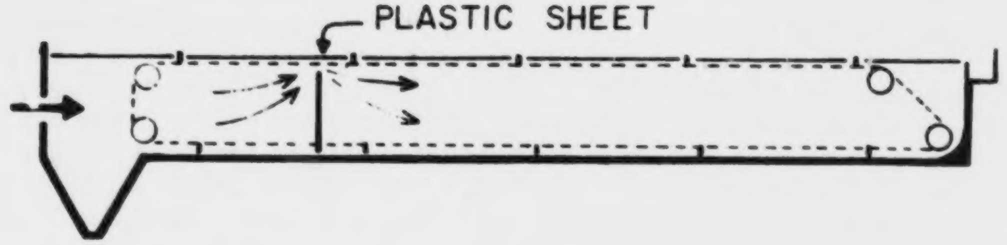
The withdrawal of sludge from secondary clarifiers for return to the aeration basin takes place at a point very near the inlet to the tank with the exception of the peripheral feed tanks. Therefore, the required area of the final clarifier is calculated using the incoming wastewater design flow and the return sludge flow rate is not included. The overflow rate for a clarifier may be calculated based on the sludge volume surface load and desired effluent suspended solids concentration. For example, for an effluent suspended solids concentration of 25 mg/l the sludge volume surface load is five gal/sq ft-hr (120 gal/sq ft-day) based on the curve in Figure 3-19. At a mixed liquor suspended solids concentration of 3000 mg/l and a sludge volume index of 100 ml/g, the volume occupied by the settled sludge after 30 minutes of settling is 30 percent of the initial volume. This calculation is shown in Equation 3-19.

$$3000 \text{ mg/l} \; \frac{100 \text{ ml}}{\text{gm}} \times \frac{\text{gm}}{1000 \text{ mg}} \times \frac{\text{L}}{1000 \text{ ml}} = 0.3 \qquad 3\text{-}19$$

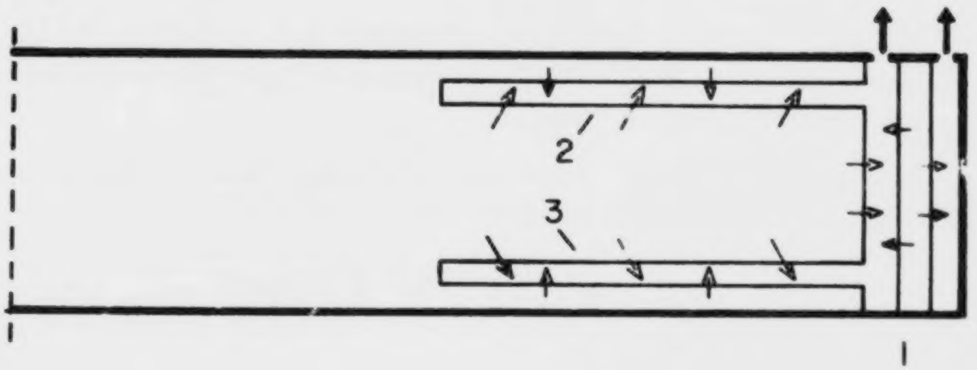
Therefore, the required overflow rate is equal to 400 gal/sq ft-day as shown by the calculation in Equation 3-20.

$$\frac{120 \text{ gal/sq ft-day}}{0.3} = 400 \text{ gal/sq ft-day} \qquad 3\text{-}20$$

NORDHORN


PLASTIC SHEET

BADEN



ERGOLZ

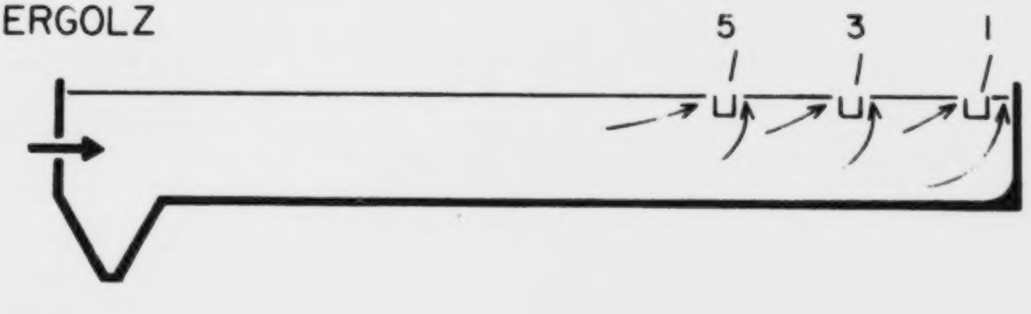


FIG.3-20

IMPROVEMENTS OF FINAL CLARIFIERS

The calculated overflow rate is somewhat lower than that frequently used for design valves of 600 to 800 gal/sq ft-day. This lower value based on the data presented by Pflanz (1968) may be attributed to the fact that in the particular secondary clarifier used by Pflanz only one effluent weir across the entire width of the rectangular basin was used at relatively low temperatures, namely between 13° - 15°C. Therefore, at newly designed plants, the overflow rate of the secondary clarifier of 600 gal/sq ft-day can be used for smaller plants up to a total of one MGD, and up to 700 gal/sq ft-day for the larger plants. The design calculations should not be based on the daily design flow but rather on the peak flow. Therefore, if the design is based on the peak hourly flow, the overflow rates for smaller plants should be 25 gal/sq ft-hour and 30 gal/sq ft-hr for the larger plants.

NITRIFICATION AND DENITRIFICATION

Nitrification and denitrification have been reported to occur in the activated sludge process. These processes remove nitrogen from the effluent from the activated sludge process.

Nitrification involves a conversion of ammonia to nitrite and nitrate. The reactions are presented in Equations 3-21 and 3-22.

$$2NH_4 + 3O_2 \xrightarrow{\underline{\text{Nitrosomonas}}} 2NO_2^- + 2H_2O + 4H^+ + \begin{matrix}\text{New}\\ \underline{\text{Nitrosomonas}}\\ \text{Cells}\end{matrix} \qquad 3\text{-}21$$

$$2NO_2 + O_2 \xrightarrow{\underline{\text{Nitrobacter}}} 2NO_2^- + \text{New } \underline{\text{Nitrobacter}} \text{ cells} \qquad 3\text{-}22$$

The conversion of ammonia to nitrite is accomplished by the microorganism Nitrosomonas, and the conversion of nitrite nitrogen to nitrate nitrogen is accomplished by Nitrobacter organisms. The rate of growth of Nitrosomonas is described by Equation 3-23 for nitrification.

$$\mu = \mu_{max} \frac{(NH_4)}{k_m + (NH_4)} \qquad 3\text{-}23$$

in which:

μ = growth rate of Nitrosomonas (day^{-1})

μ_{max} = maximum growth rate (day^{-1})

(NH_4) = concentration of ammonia (mg/l)

k_m = concentration of ammonia when $\mu = 0.5\mu_{max}$ (mg/l)

Downing (1964) reported the growth constants for Nitrosomonas at 20°C to be $\mu_{max} = 0.33$/day and $k_m = 1.0$ mg/l of ammonia nitrogen. The maximum growth rate for the Nitrobacter at 20°C reported by Downing was $\mu_{max} = 0.14$/day. These data indicate that the maximum growth of Nitrosomonas and Nitrobacter probably occur at an ammonia concentration of about 3.0 mg/l. At concentrations of ammonia above this minimum concentration, the rate of nitrification is independent of the ammonia concentration and the nitrification reactions are zero order. The growth rate constants for the nitrifying bacteria are relatively low compared to those reported for heterotrophs. For example , Schulze (1964) reported that the maximum growth rate for Escherichia coli and a glucose substrate was 18 per day. Therefore, the detention time in the activated sludge system must be greater than the generation time of the nitrifying bacteria, otherwise these bacteria would be washed out of the system. Since the nitrifying bacteria are associated with the mixed liquor the sludge age should be sufficiently long to provide active growth of the nitrifying bacteria if nitrification is to take place. On the other hand, the amount of nitrification which occurs can be calculated from the sludge age.

The maximum growth rate of the nitrifying bacteria is markedly affected by changes in temperature. The result of investigations at the Water Pollution Laboratory (Anon., 1967) indicate that the rate constant increases about seven percent per degree centigrade. The effect of temperature on the growth of nitrifying bacteria is presented in Figure 3-21 Nitrification will take place if the reciprocal sludge age (1/G) is lower than the maximum growth rate of the nitrifying bacteria.

Aerobic conditions must be maintained for nitrification. Downing (1964) and Wuhrmann (1964) reported that a minimum concentration of dissolved oxygen of one mg/l is required in order to achieve nitrification. The maximum rate of nitrification requires dissolved oxygen concentrations of more than 200 mg/l.

Nitrification will occur to some extent in activated sludge treatment plants operated at organic loading rates of about 0.25 lb BOD/lb MLSS-day or lower. If nitrification does occur, a higher oxygen uptake rate will be exerted than that calculated for the removal of carbonaceous matter only. Therefore, nitrification should be included in calculating oxygen requirements.

Partial nitrification is possible in the contact stabilization process, since the sludge age is relatively high in this process. However, since the contact time is relatively short, only a partial nitrification is possible. In the

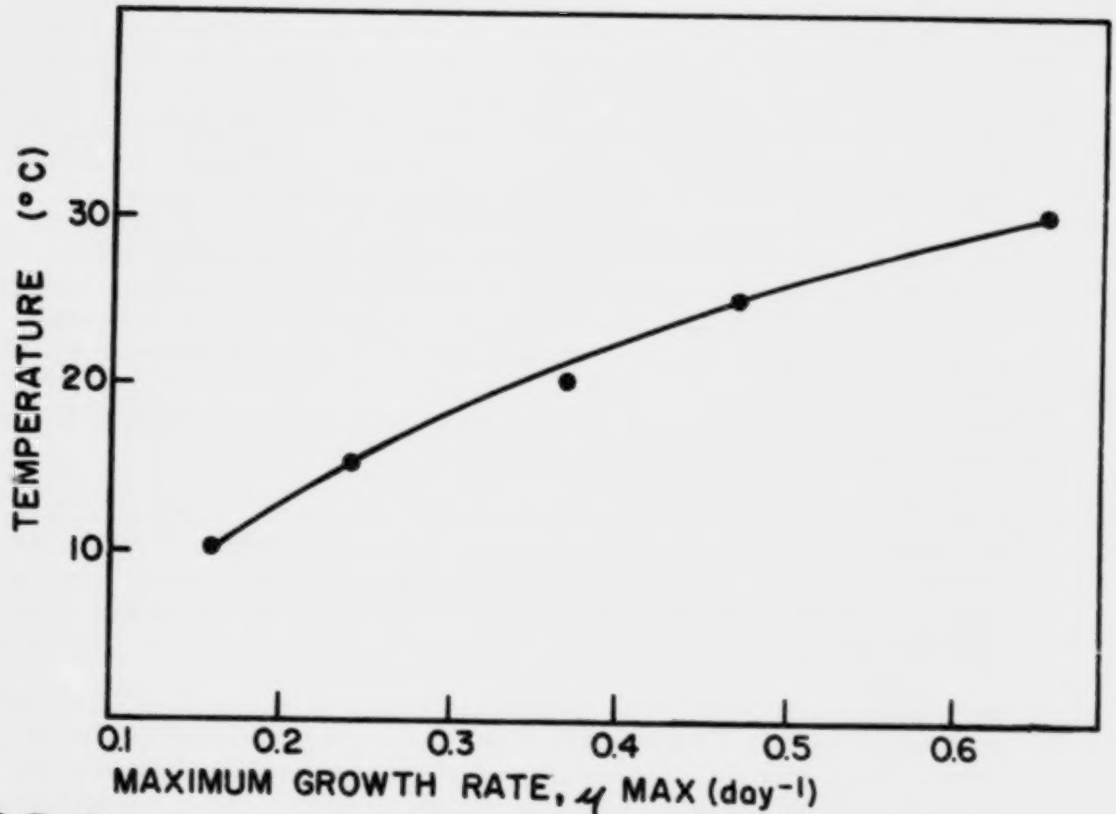


FIG. 3-21
TEMPERATURE EFFECTS ON GROWTH RATE OF NITRIFYING BACTERIA (Downing 1964)

stabilization tanks where the concentration of nitrifying bacteria is high, the concentration ammonia is the limiting factor. The ammonia removal in the contact stabilization process will be increased at higher return sludge flow rates and lower contact times.

Denitrification can take place once the ammonia has been converted to the nitrite or nitrate form. If the dissolved oxygen concentration in the mixed liquor is at a concentration less than one mg/l heterotrophic bacteria can utilize the oxygen associated with the nitrites and nitrates for metabolism, and the nitrogen is released as nitrogen gas. A denitrified effluent can be obtained by placing a denitrification tank between the aeration basin and the final clarifier. The denitrification tank should be equipped with a mixing device to maintain the solids in suspension but not to introduce any oxygen. The time required for total denitrification can be estimated from the initial dissolved oxygen concentration in the mixed liquor and the quantity of oxygen present in the form of nitrites or nitrates as well as the oxygen uptake rate of the mixed liquor. Wuhrmann (1964) indicated that the time required for denitrification can be calculated from the mass balance which is presented in Equation 3-24:

$$VR_N = Q\,(C_o + C_N) \qquad 3\text{-}24$$

$$t_D = \frac{V}{Q} = \frac{C_o + C_N}{R_N} \qquad 3\text{-}25$$

in which:

t_D = detention time of the mixed liquor in the denitrification basin (hrs)

C_o = dissolved oxygen concentration in the mixed liquor entering the denitrification basin (mg/l)

C_N = oxygen in the form of nitrite and nitrates (mg/l)

R_N = oxygen uptake rate of the mixed liquor in denitrification basin (mg/l-hr)

The oxygen uptake rate in the denitrification basin is very similar to the endogenous rate. If nitrification takes place in the aeration tank an no denitrification is provided, partial denitrification may take place in the final clarifier and the released nitrogen gas may cause a portion of the settled sludge to float. Therefore baffles should be located near the effluent weirs in the final clarifiers to reduce the carryover of suspended solids into the effluent.

PROCESS CONTROL

The practical control of the activated sludge process involves:

(a). control of the concentration of dissolved oxygen in the aeration tank, or

(b) control of the mixed liquor suspended solids concentration.

The dissolved oxygen concentration in the aeration basin should usually be maintained between 1.0 and 2.0 mg/l. The amount of oxygen transferred in the aeration basin therefore should be capable of satisfying variations in the oxygen uptake rate. The automatic control of the oxygen transfer to the system may be realized by monitoring the concentration of dissolved oxygen at the point in the aeration tank where the oxygen uptake rate is at a maximum. If the concentration of dissolved oxygen in the aeration tank falls below a minimum desirable level the aeration equipment can be called upon to provide additional oxygen. On the other hand, if the dissolved oxygen concentration in the aeration basin exceeds the desired level, the amount of oxygen introduced can be reduced. The savings in power costs over a number of years of operation would probably be sufficient to cover the capital cost of the automatic control system. A single oxygen probe would be required for a completely mixed aeration tank or the tank in a step aeration system, since the oxygen uptake rate should be the same throughout the entire volume of the tank. In longitudinal spiral flow tanks generally used for the conventional activated sludge process, the oxygen probe should be installed near the wastewater inlet since it is at this point that the oxygen uptake rate is the highest.

The concentration of mixed liquor suspended solids in the aeration tank can be decreased by increasing the quantity of sludge wasted. An increase in the mixed liquor suspended solids concentration is practically impossible; therefore, the quantity of sludge wasted should be carefully controlled. Generally, continuous wasting at a predetermined rate is preferred to the practice of wasting sludge periodically.

Guarino and Carpenter (1970) indicated that the control of the loading rate is possible, by storing return sludge to provide the necessary sludge to maintain a constant loading rate. This procedure is essential since the incoming BOD is very difficult to control. Therefore, by controlling the concentration of mixed liquor suspended solids. the loading rate can be maintained relatively constant. Sludge is pumped to storage tanks during the night time hours when the loading rates are relatively low and introduced into the system during the day time when the organic loading reaches its peak. This system would provide savings in power required for accommodating the oxygen uptake rate.

All municipal wastewater treatment plants operate under transient loading conditions. The ratio of the daily peak to the minimum flow is in a range of 3-1 to 10-1 or more. In most treatment plants, the effluent BOD is practically not affected by the fluctuating transient conditions provided the organic loading was equal to or less than 0.5 lb BOD/lb MLSS-day. Observations at treatment plants indicate that the biological processes require some time for recovery. Generally the recovery occurs during the night time hours when the organic load to the plant is low or on the weekends when loadings are also lower than during the work week. The recovery time could be required for stabilization of the biological solids and/or for the biodegradation of other particulate material which accumulate in the mixed liquor.

DESIGN FACTORS

Design and layout of an activated sludge plant for the treatment of municipal wastewater is based on the following design data:

(a) wastewater flow (MGD) daily peak wastewater flow (1000 gal/hr)

(b) average influent BOD (mg/l)

(c) average influent suspended solids (mg/l)

The basic data should also include any industrial waste and information relating to type, source of wastewater, and quantity of discharge. This information is required to minimize the possibility of shock loadings to the activated sludge system. If more than 1/2 the total BOD load to the plant is contributed by a single industry, it is conceivable that laboratory or pilot-scale treatability studies should be performed to effectively select the proper design parameters.

A number of other factors must be considered in the design of a biological treatment plant since other variables might affect the performance and operation of the overall plant. The need for primary sedimentation prior to biological treatment is sometimes questioned. The characteristics of the wastewater and the manner in which the solids are to be ultimately disposed of should be considered carefully prior to deciding on whether a primary clarifier is required. The removal of suspended solids in the primary sedimentation tank would minimize the load on the biological treatment process since some of the biodegradable solids would be removed and the oxygen uptake requirements would be considerably lower. In this case, the primary solids would more than likely be treated in an anaerobic digestion system prior to disposal on the sludge drying beds. However, it is also possible for the primary solids to be mixed with the waste activated sludge and treated by aerobic digestion provided the quantity of waste activated sludge is at least equal to the quantity of primary sludge requiring disposal.

Generally, the final decision is based on a cost estimate. However, other factors should also be included before a final decision. Primary sedimentation is preferable in those cases where the incoming suspended solids are mainly inorganic such as might be found in a treatment plant handling combined wastewater. Primary clarification also permits the removal of floating materials. Therefore, if primary clarification is not installed prior to the aeration tank, it is essential that the secondary clarifiers be equipped with skimming equipment.

The layout of the biological treatment plant is also important. Any layout of the processes should provide for flexible operation under a variety of conditions. Most municipal waste treatment plants have a number of aeration basins and final clarifiers operating in parallel. This requirement is generally established by the regulatory agency in the particular state or region. However, pipes and channels connecting the various units should be so designed that taking any one unit process out of operation does not affect the use of the remaining unit processes. For example, it should be possible to take an aeration tank out of service and operate all clarifiers by distributing the incoming wastewater, and return sludge to the remaining aeration units. On the other hand, if a final clarifier is out of operation, it should be possible to continue operating the plant with all unit aeration tanks being utilized. With this type of design and plant layout a single blower station and a single sludge return pumping station could be included in the design. Some additional piping might be required to maintain this flexibility, but the cost of the piping is recovered in terms of the ability of having continuous operation of the plant.

An estimate of the ultimate organic load and wastewater flow must be determined at the time of the initial design, although it is extremely difficult to estimate what might happen 25 to 50 years after the plant is in operation. The future program of operation of a treatment plant is helpful in the design of the various components so that the plant can be expanded as organic and hydraulic loads increase. The initial construction phase of the treatment plant should be designed for ease of expansion.

DESIGN FORMULATIONS AND EXAMPLES

The performance of the activated sludge process in terms of BOD concentration in the effluent can be used for design of the process. The relationship between the effluent BOD and the BOD loading was presented in Figure 3-7. The design BOD effluent concentration is determined by taking the maximum effluent BOD at a given organic loading and increasing that value by about 50 percent. This increase is necessary because the data represent average values and the standard deviation for these data was about 0.5 of the average. The effluent suspended solids concentration was derived from the relationship developed in Figure 3-9 which relates the effluent suspended solids concentration and the sludge age.

The basic design data are listed in Table 3-1. Columns one and two represent the particular parameter in question and the dimensions of parameter, respectively. Column 3 represents the calculations for the extended aeration process. The effluent qualities for the activated sludge process at various loadings are summarized in Columns 4, 5, 6, 7, and 8, respectively. Two sets of calculations for the oxygen uptake rate and the oxygen transfer are presented for the activated sludge process for the loading rate of 0.25 lb BOD/lb MLSS-day. One set of calculations represents the oxygen requirements for biodegradation of only carbonaceous material while the other set of calculations includes the oxygen requirements for nitrification of ten mg/l in addition to the carbonaceous oxygen requirements.

The incoming wastewater contains 150 mg/l of BOD and 150 mg/l of suspended solids. These design calculations can be applied to wastewaters which have an incoming BOD which ranges from 100 to 200 mg/l provided the ratio of incoming BOD to incoming SS is 1:1. The information presented in the various lines in Table 3-1 will be discussed below where discussion is necessary.

Line 6 Design MLSS:

3000 mg/l has been assumed for all processes and process modifications; however, in practice a MLSS concentration up to 4000 mg/l can be applied.

Line 7 BOD Volume Load:

$$(\text{lb}/1000\ \text{gal/day}) = \frac{\text{BOD Load } 8.34}{1000}$$

$$\frac{\text{lb BOD}}{\text{lb MLSS-day}} \quad \frac{\text{mg/l}}{10^6} \quad \frac{8.34\ \text{lb}}{\text{gal}}$$

Line 8 Excess Sludge

$$\frac{1}{G} = \frac{0.6\,(S_r + X_o)}{Xt} - 0.075$$

in this example, $S_o = X_o$ or $S_o/X_o = 1$

TABLE 3-1

ACTIVATED SLUDGE PROCESS DESIGN CHART

		Dimensions	Extended Aeration	Activated Sludge				
1.	Design BOD Effluent	mg/l	(15)**	20		25	35	50
2.	Average BOD Effluent	mg/l	(10)**	10		12	15	22
3.	Design Suspended Solids Effluent	mg/l	(-65)*	25		30	35	45
4.	Average Suspended Solids Effluent	mg/l	(-15)**	15		16	18	22
5.	Design BOD Load	lb/lb/day	0.065	0.25		0.50	1.0	2.0
6.	Design MLSS	mg/l	3000	3000		3000	3000	3000
7.	BOD Volume Load	lb/10^3 gal/day	1.60	6.25		12.5	25	50
8.	Excess Sludge	lb/lb MLSS/day	0.03	0.22		0.52	1.12	2.32
9.	Oxygen Uptake Rate	lb/lb MLSS/day	0.185	0.370	0.275	0.40	0.65	1.15
10.	per gallon tank	lb/10^3 gal/day	4.40	9.30	6.90	10.0	16.20	28.75
11.	Oxygen Concentration, ml	mg/l	1	2	2	2	2	2
12.	Oxygen Transfer	lb/10^3 gal/day	5.20	12.20	9.10	13.20	21.50	38.0
13.	Airflow/gallon tank	cu ft/gal/day	3.60	8.60	6.40	9.30	15.00	26.5
14.	Airflow/lb BOD Appl.	cu ft/lb BOD	2250	1360	1000	740	600	550
15.	Peak (24/18)	cu ft/lb BOD	2250	1820	1350	1000	800	733
16.	Design Airflow	cu ft/lb BOD	2800	2400	2000	1500	1200	1100
17.	Detention Time	hours	(19)	(4.80) 3.85		(2.40) 1.90	(1.20) 0.95	(0.60) 0.48
18.	Airflow/Gallon Sewage	cu ft/gal	(2.90)	(1.70) 1.37	(1.27) 1.02	(0.93) 0.74	(0.75) 0.60	(0.69) 0.55
19.	Waste Sludge/gallons	lb/10^3 gal	(0.42)**	(0.97) 0.73		(1.15) 0.92	(1.20) 0.96	(1.25) 1.00

* If no sludge is wasted (all sludge leaves through effluent)

**If sludge is wasted

therefore

$$\frac{1}{G} = 0.6\ (2)\ \text{BOD Load} \quad - 0.075$$

Line 9 Oxygen Uptake Rate:

(lb O_2/lb MLSS - day)

$$\frac{R}{X} = a'\ \left(\frac{S_o - S_e}{Xt}\right) + b' \quad \text{(eq 3-3)}$$

$$\frac{R}{X} = 0.5\ (\text{BOD Load}) + 0.15$$

but for nitrification at BOD Loads of 0.25 lb/BOD/lb MLSS - day.

Oxygen uptake for nitrification is:

$$\frac{R}{X} = 4.6\ \frac{(10)(24)}{3000\ (3.85)} = 0.095$$

Total uptake rate including nitrification

$$(0.5\ (0.25) + 0.15) + 0.095 = \frac{0.370\ \text{lb } O_2}{\text{lb MLSS-day}}$$

Line 10 Oxygen Uptake Rate per 1000 gal of Tank Volume:

$$\frac{R}{X} = X\frac{(8.34)}{1000}\ \left(\frac{\text{lb } O_2 \quad \text{MLSS mg/l}\ \ 8.34\ \text{lb}}{\text{lb MLSS -day}\ \ 10^6\ \text{gal.}}\right)$$

Line 11 Dissolved Oxygen Concentration in Mixed Liquor:

mg/l

Line 12 Oxygen Transfer:

lb O_2/1000 gal - day

$$\text{Oxygen Uptake}\ \left(\frac{C_{S_M}}{C_{S_M} - C}\right)$$

C_S at standard conditions: 8.0 mg/l

σ 0.9

Compensation for 13 foot depth of air release from Figure 3-13

$C_{S_M} / C_S = 1.16$

$$\therefore \; C_{S_M} = C_S \, \sigma \, \frac{C_{S_M}}{C_S}$$

$C_{S_M} = 8.0\ (0.9)\ 1.16 = 8.3$ mg/l

Line 13 Airflow Rate:

(Cu ft/gal - day)

Specific oxygen transfer = 0.11 lb O_2/1000 cu ft/ft at 13 foot depth of air release. 0.11 (13) = 1.43 lb O_2/1000 cu ft of air.

$$\frac{\text{Oxygen Transfer/1000 gal - day}}{1.43 \text{ lb } O_2\text{/1000 cu ft air}}$$

Line 14 Average Airflow per lb BOD Applied:

$$\frac{\text{Oxygen uptake rate}}{\text{BOD Load (1.43)}} \quad \frac{C_{S_M}}{C_{S_M} - C} \quad \cdot \; 1000$$

Line 15 Airflow at Peak BOD:

(cu ft/lb BOD)

Peak hour = 1/18 of total daily

Peak airflow = $\frac{24}{18}$ Average Airflow

Line 16 Design Airflow:

(cu ft/lb BOD)

The design airflow is required for the layout of the blowers and includes a factor of safety of 1.25 for extended aeration plants and 1.50 for the activated sludge plants. Design Airflow = peak airflow (1.25) or Peak airflow (1.50).

Line 17 Detention Time:

(hours)

$$\frac{S_o (24)(8.34)}{(\text{BOD Volume load})\ 1000}$$

The numbers in ( ) in Table 3-1 are based on $S_o = 150$ mg/l.

The other values of detention time are based on $S_o = 120$ mg/l

Line 18 Airflow per Gallon of Wastewater:

(cu ft/gal)

$$\frac{(\text{cu ft/gal of tank volume})\ (\text{Detention time})}{24 \text{ hrs/day}}$$

$Q = V/t \therefore t/V = 1/Q$

Line 19 Waste Sludge:

(lb sludge/1000 gallons wastewater)

$$\frac{0.6\,(S_o + X_o) - (0.075\,\frac{Xt}{24}) - X_e}{X_o}$$

The data in Table 3-1 will be used for this design example for an activated sludge plant with primary clarification.

Sewage flow	20 MGD
Peak Hour	26.5 MGD
Primary Effluent	120 mg/l BOD, 120 mg/l SS

Total BOD after primary clarification is:

20 MGD (120 mg/l) 8.34 lb/gal = 20,000 BOD/day

Design for effluent BOD	20 mg/l
Design load	0.25 lb BOD/lb MLSS - day
Design MLSS	3000 mg/l
BOD volume load (from Table 3-1)	6.25 lb BOD/1000 gal/day

Aeration tank volume:

$$\frac{(20,000 \text{ lb BOD/day}) \ 1000 \text{ gal}}{6.25 \text{ lb BOD/day}} = 3.2 \times 10^6 \text{ gallons}$$

Detention Time:

$$\frac{3.2 \times 10^6 \text{ gal.}}{20 \times 10^6 \frac{\text{gal}}{\text{day}}} \quad \frac{24 \text{ hours}}{\text{day}} = 3.84 \text{ hours}$$

Excess sludge = 0.22 lb/lb MLSS - day

Waste Sludge:

$$\text{Total MLSS} = 3.2 \times 10^6 \text{ gal} \left(\frac{3000 \text{ mg/l}}{10^6}\right)\left(\frac{8.34 \text{ lb}}{\text{gal}}\right) = 80,000 \text{ lb}$$

$$\frac{0.22 \text{ lb Solids}}{\text{lb MLSS - day}} \ (80,000 \text{ lb MLSS}) = 17,600 \text{ lb/day}$$

Return sludge concentration at a return sludge flow rate equal to the flow rate of the incoming wastewater and MLSS concentration, X = 3000 mg/l.

$$Q_{RS}X_{RS} = X(Q + Q_{RS})$$

$$X_{RS} = X\frac{Q + Q_{RS}}{Q_{RS}}$$

$$X_{RS} = 3000\ (2/1)$$

$$X_{RS} = 6000 \text{ mg/l}$$

Flow Rate of Waste Sludge

$$\frac{17,600 \text{ lb/day} \times 10^6}{(6,000 \text{ mg/l})\ 8.34 \text{ lb/gal}} = 352,000 \text{ gallons/day}$$

Oxygen Transfer (Average): 9.10 lb O_2/1000 gal/day

Total oxygen transfer per day

$$\frac{9.10 \text{ lb } O_2}{1000 \text{ gal-day}} \ 3.2 \times 10^6 \text{ gal} = 29,000 \text{ lb } O_2\text{/day}$$

At peak hour the oxygen transfer rate is:

$$\frac{29{,}000}{} \ \frac{\text{lb O}_2}{\text{day}} \ \frac{\text{day}}{24\text{ hr}} \left(\frac{24}{18}\right) = 1{,}600 \text{ lb/hour}$$

Including the Safety factor for design, the peak oxygen transfer rate is

$$1{,}600 \ \frac{\text{lb}}{\text{hr}}(1.5) = 2{,}400 \text{ lb/hour}$$

Aeration Equipment

Cone type aerators could be used and each aerator has an oxygenation capacity of 240 lb O_2/hour; therefore, the number of aerators required is

$$\frac{2400}{240} = 10 \text{ aerators}$$

The variation in oxygen transfer can be accomplished by changing the depth of immersion by the ratio of 1:3; therefore, the minimum oxygen transfer rate is about 800 lb O_2/hour.

Diffused Aeration System

Specific oxygen transfer rates = 0.11 lb O_2/1000 cu ft per foot of depth of air release

Therefore, at a diffuser depth of 13 feet, the oxygen transfer rate is

$$\frac{0.11 \text{ lb O}_2}{1000 \text{ cu ft-ft}} \ 13 \text{ ft} = 1.43 \text{ lb O}_2/1000 \text{ cu ft}$$

The design airflow rate is

$$\frac{2400 \text{ lb/hr } 1000 \text{ cu ft}}{(1.43 \text{ lb}) \ 60 \text{ min/hr}} = 28{,}000 \text{ cfm}$$

The design airflow per lb BOD is

$$\frac{28{,}000 \text{ cu ft/min}}{20{,}000 \text{ lb BOD/day}} \left(\frac{1440 \text{ min}}{\text{day}}\right) = 2000 \text{ cu ft/lb BOD}$$

The airflow at the daily peak is

$$\frac{2000 \text{ cu ft/lb BOD}}{1.5} = 1{,}350 \text{ cu ft/lb BOD}$$

AERATED LAGOON PROCESS

PROCESS DESCRIPTION

Aerated lagoons are ponds in which diffused air or mechanical aeration systems are installed. Detention times of less than two days to more than 30 days have been employed.

An aerated lagoon is generally visualized as a completely mixed tank. The biodegradation process is controlled only by maintaining aerobic conditions in at least part of the basin. The microbial population is controlled by the concentration of substrate entering the lagoon and the detention time. However, there is almost no control of the influent BOD after the aerated lagoon system is put into operation. The power introduced into the aerated lagoon for aeration and mixing establishes the extent to which sedimentation of the solids takes place in the ponds. At high power levels almost all of the solids are maintained in suspension and little settling occurs. These systems are completely mixed and can be considered to be aerobic lagoons. At lower power levels there is some settling of solids to the bottom of the pond where the accumulated biodegradable solids undergo anaerobic decomposition. These lagoons are facultative with an aerobic zone overlying the anaerobic area.

In general, the suspended solids concentration in the effluent of aerated lagoons is considerably higher than for the activated sludge process. The effluent suspended solids concentration can be reduced somewhat by baffling a segment of the pond to permit a quiescent zone where the suspended solids may settle. An unmix pond may be installed following the aerated lagoon to permit the removal of the settleable suspended solids. Algae may develop and the effluent suspended solids concentration may not change but the composition will be different.

PERFORMANCE OF THE AERATED LAGOON PROCESS

The mathematical model which is used to describe the kinetics of the aerated lagoon process does not include the active mixed liquid suspended solids since the concentration of suspended solids can be controlled only by controlling the power level. The equation may be written as:

$$S = \frac{1}{akt} + \frac{b}{ak} \qquad 4\text{-}1$$

Equation 4-1 indicates that the substrate concentration in the effluent BOD decreases with increasing detention time and effluent BOD is independent of the influent BOD. The concentration of active mixed liquor suspended solids in an aerated lagoon operated at a given detention time and power level will increase as the influent BOD increases. Therefore, the effluent BOD concentration remains relatively constant. The substrate concentration in the lagoon, S, includes both soluble and insoluble organic material. However, only the BOD of the soluble fraction is measured. In general, it can be assumed therefore that the insoluble fraction, if it is measured as BOD is relatively low since the concentration of microorganisms in the lagoon is much higher than that which is found in the BOD bottle.

Equation 4-1 can be evaluated graphically by plotting the substrate concentration, S, versus the reciprocal of time $(\frac{1}{t})$. The straight line through the data will have a slope of 1/ak and the intercept will be equal to b/ak. However, there is some question whether the total or soluble BOD should be used. One relationship will be developed for only the soluble BOD in which it is assumed that the fraction of the total BOD removed is the same as the fraction of the soluble BOD removed or

$$S_e^*/S_o^* = S_e/S_o \qquad 4\text{-}2$$

The results of laboratory-scale aerated lagoon experiments at very short detention times are presented in Figure 4-1. There is some scatter among the data points; however, the slope of the line through the data is 7.0 (mg/l of BOD) (days). Therefore, the product ka = 1/7.0 = 0.143. For municipal wastewater a = 0.65; therefore, k = 0.22 $(\text{day}^{-1})\ (\text{mg/l})^{-1}$. The intercept in Figure 4-1 is difficult to determine. However, by assuming an endogenous respiration rate, b= 0.15 per day, the intercept can be calculated as 0.15/7.0 = 0.02 mg/l of BOD. Comparison of this value to the measured BOD and considering the scatter of the data, the line describing the data could have been drawn through the origin as well.

The rate constant for aerated lagoon systems can also be calculated by Equation 4-3.

$$\frac{S_o - S}{X_a t} = kS \qquad 4\text{-}3$$

The terms can be rearranged and the rate of substrate removal can be expressed as:

$$\frac{S_o - S}{t} = kX_aS \qquad 4\text{-}4$$

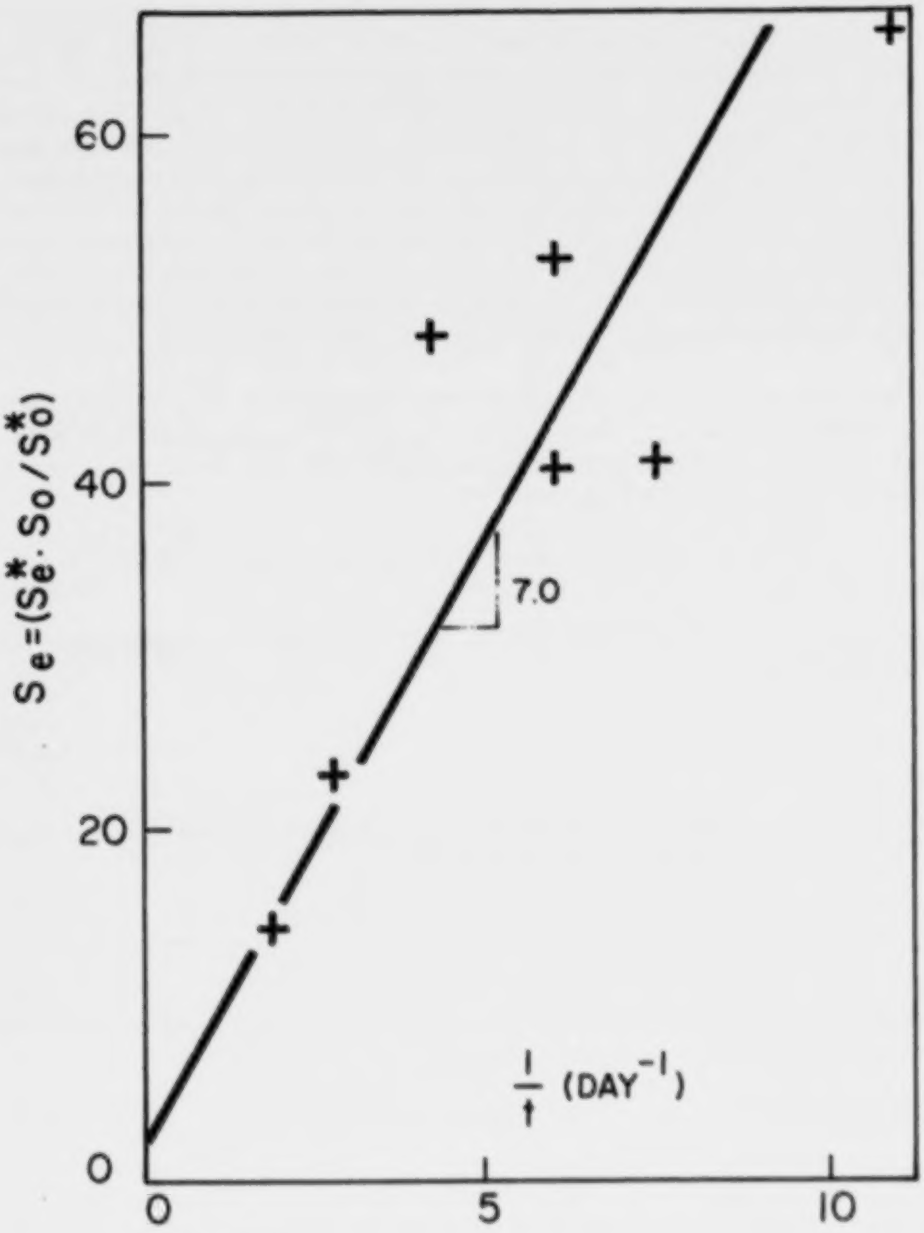


FIG. 4-1
RATE CONSTANT - LABORATORY SCALE
AERATED LAGOONS

The terms k and X_a have frequently been combined as a single constant $K = (kX_a)$. This equation has been used to correlate aerated lagoon data. A plot of the removal rate $(S_o - S/t)$ versus the effluent BOD, S, generally results in a straight line the slope of which is $K = (kX_a)$. Typical data are plotted in Figure 4-2 and the straight line describing these data has a slope of 8.0. The rate constant was determined in these experimental studies and k = 0.21; therefore, the active mixed liquids suspended solids concentration $X_a = 8/0.21$ or approximately 40 mg/l. The data plotted in Figure 4-2 are the results of analyses of composite samples collected from laboratory units which were loaded at a constant rate. These laboratory units were operated under nearly steady state conditions.

In the full-scale operation the influent wastewater flow varies; therefore, the detention times in the aerated lagoon and the influent BOD also vary. The equation for the unsteady state conditions can be derived from a mass balance expressed as Equation 4-5.

$$QS_o\Delta t - QS_e\Delta t = kX_aVS_e\Delta t + V\Delta S \qquad 4\text{-}5$$

The mixed liquid suspended solids concentration (X_a) can be expressed as Equation 4-6.

$$X_a = \frac{a(S_o - S_e)}{1 + bt} \qquad 4\text{-}6$$

The right hand side of Equation 4-6 can be used in Equation 4-5 in place of X_a and rewritten as Equation 4-7.

$$QS_o\Delta t = QS_e\Delta t + kVS_e\Delta t \left(\frac{a(S_o - S_e)}{1 + bt} \right) + V\Delta S \qquad 4\text{-}7$$

Equation 4-7 can be rearranged as Equation 4-8 and used to calculate the change in substrate concentrations.

$$\Delta S = t(S_o - S_e)\left[\frac{1}{t} - \frac{kaS_e}{1 + bt}\right] \qquad 4\text{-}8$$

Equation 4-8 was used to calculate the variations in the effluent BOD caused by transient loadings to the aerated lagoon system. The influent BOD varied from 60 mg/l during the 12-hour nighttime period up to 120 mg/l during the 12-hour daytime period. The theoretical detention times in the aerated lagoon during these loadings were 1.2 days and 0.8 days, respectively, during the night and the day hours. The average BOD for the 24-hour period was 100 mg/l and the theoretical detention time was 1.0 days. The transient conditions are illustrated in Figure 4-3. The

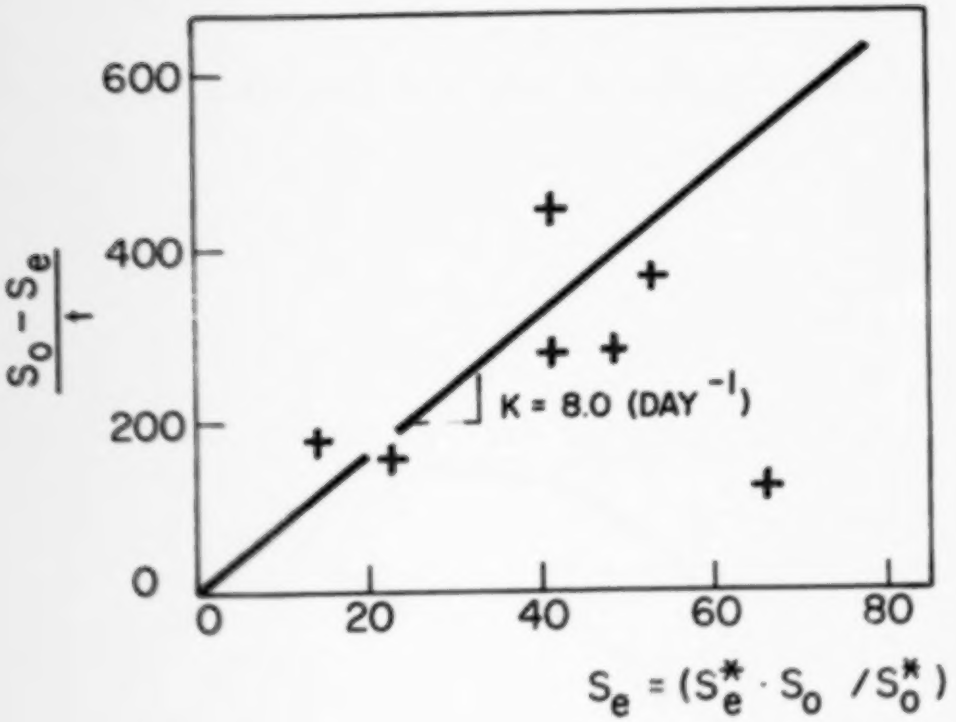


FIG.4-2
LAB SCALE AERATED LAGOONS —
EVALUATION OF THE REMOVAL
RATE CONSTANT, "K"

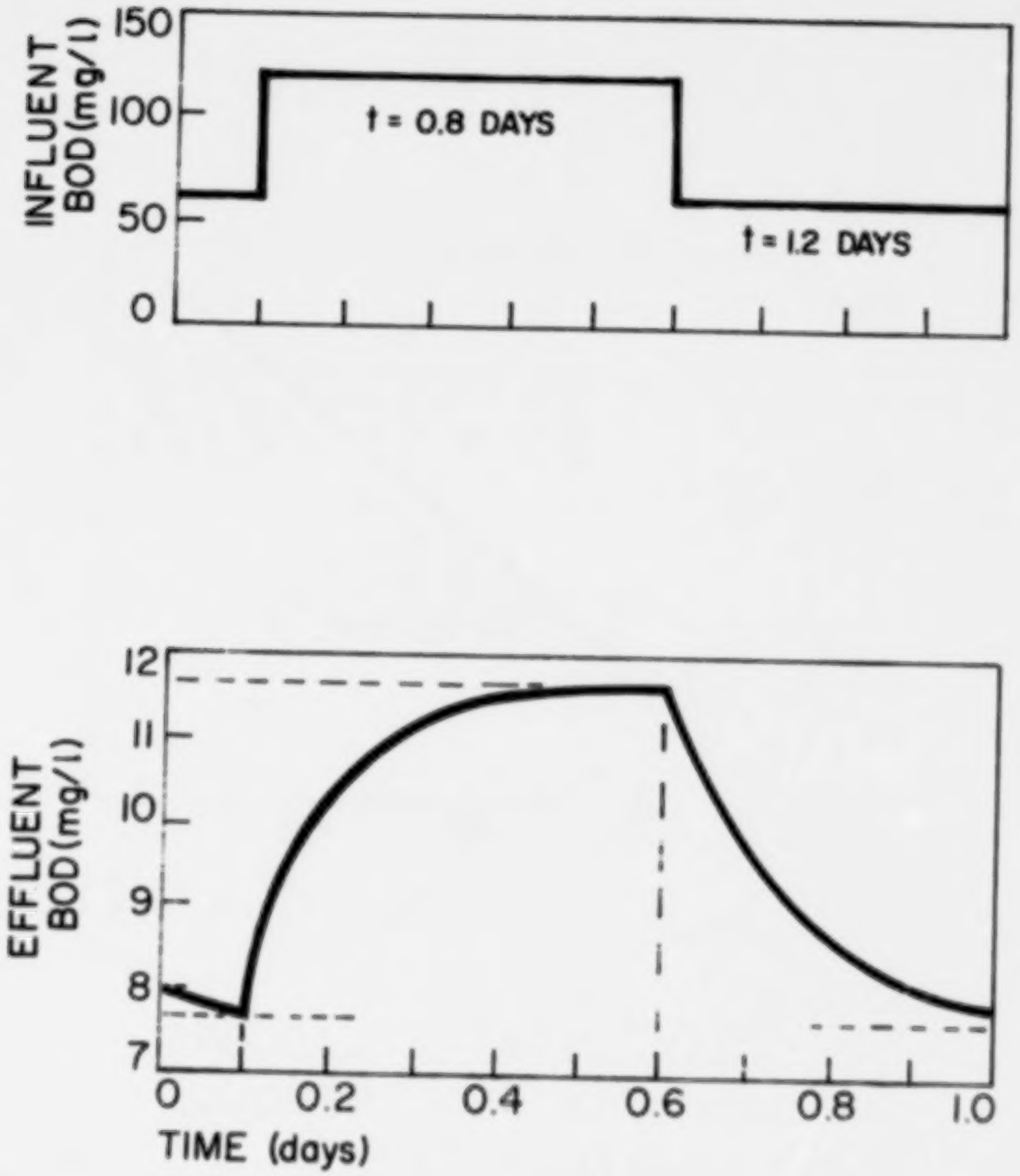


FIG 4-3
AERATED LAGOON TRANSIENT CONDITIONS

effluent BOD concentrations in Figure 4-3 varied from a minimum concentration of 7.6 mg/l at the end of a nighttime period and increased to a maximum concentration at the end of the 12-hour daylight period of 11.6 mg/l. As an influent BOD concentration decreased during the nighttime hours, the effluent BOD concentration also decreased and reached a minimum concentration at the early morning at a value of 7.6 mg/l. However, the calculated effluent BOD using a one-day detention time and the same rate constant was 8.85 mg/l. This value compares closely to the effluent BOD of 9.1 mg/l based on the composite samples taken throughout the day. It should be pointed out at this time that the variations in the effluent BOD concentration is caused primarily by the changes in the detention time in the aerated lagoon system. The effects of the changes of the influent BOD concentration are slight and only affect the slope of the curve. Therefore, decreasing the detention time accompanied by an increase in the influent BOD results in a high effluent BOD under steady state conditions.

The removal of soluble BOD is of interest for design purposes. The results of laboratory-scale and full-scale experiments at the Williamson Creek Wastewater Treatment Plant in Austin, Texas are plotted in Figure 4-4. The detention time in the laboratory-scale units ranged from 0.09 to 0.54 days (2.2 hours to 12.6 hours) while the detention time in the lagoons at the plant varied from 0.7 to 1.0 days, and 3 - 8 days for the C plant and B plant, respectively. The data in Figure 4-4 indicate that the laboratory results cluster at one part of the graph and the full-scale data cluster in another area of the graph. The line through the laboratory-scale data has a slope of minus 1 which indicates that the effluent BOD concentration is proportional to reciprocal time (1/t). These results confirm the theoretical considerations which were discussed earlier. A line through the points which cluster around the one-day detention time for the field-scale aerated lagoons also has a slope of minus 1. The detention time does not seem to affect the effluent BOD beyond the detention time of 2.5 days. At these long detention times the effluent BOD was relatively constant and the soluble BOD ranged from three to seven mg/l. The data indicate that the soluble BOD removal rate constant in the full-scale process was less than that observed for the laboratory-scale experiments.

It is doubtful that the effluent soluble BOD concentration could ever be reduced to below four to five mg/l even at longer detention times. The intersection of the line describing the laboratory data and a horizontal line at four to five mg/l effluent soluble BOD occurs at a detention time of about 0.6 days. This detention time is approximately equal to that at which the maximum active mixed suspended solids was reported in the laboratory-scale lagoons. Increasing the detention time beyond the point at which the maximum active mixed liquids suspended solids concentration occurs will not reduce the

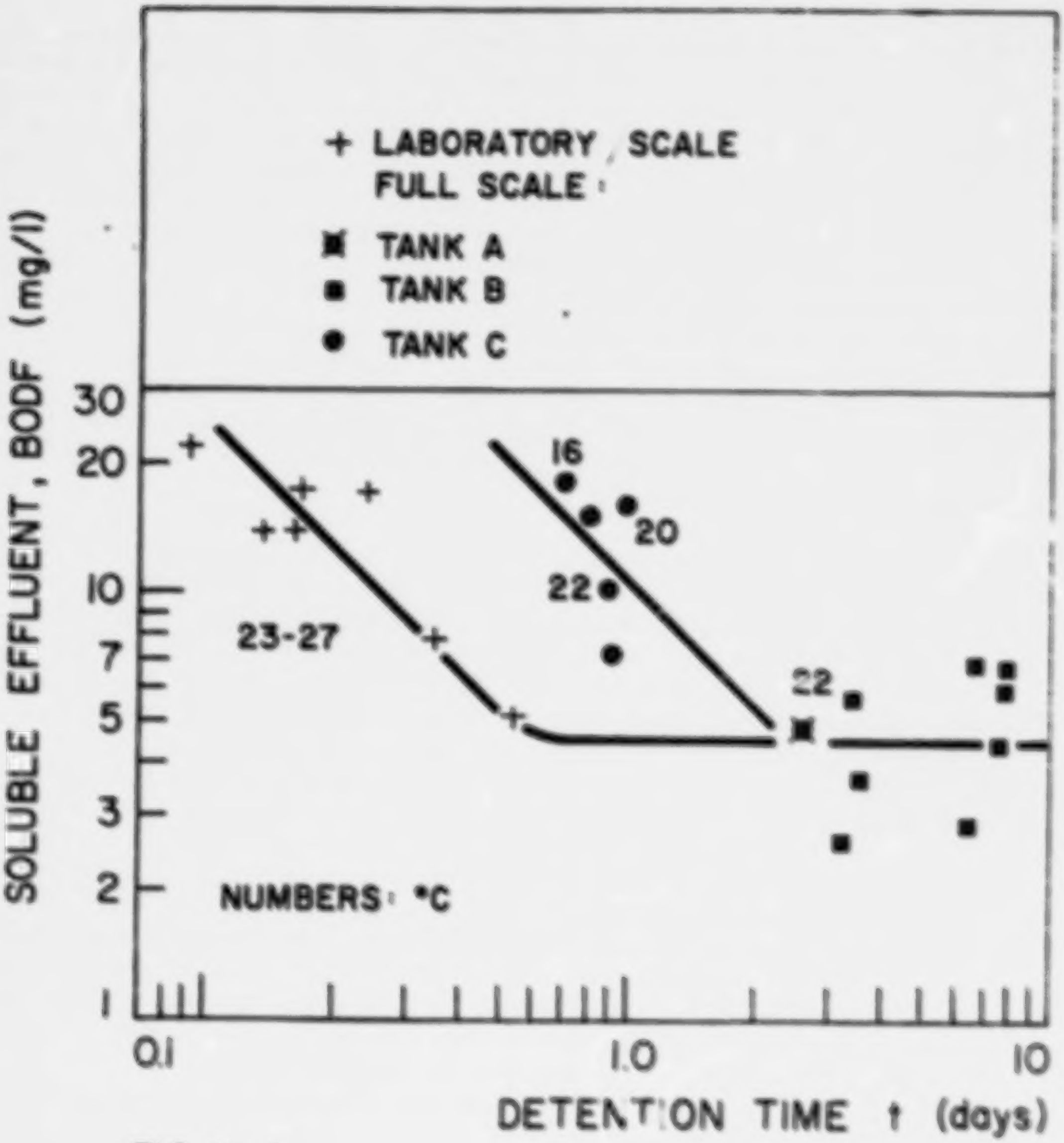


FIG.4-4

SOLUBLE EFFLUENT - BOD VERSUS DETENTION TIME effluent BOD concentration since at detention times beyond this point, the active mixed liquids suspended solids concentration decreases and less soluble BOD will be removed. In full-scale systems, this breakpoint can be assumed to occur at a detention time of about 2.0 days. The difference in the rate constants can be attributed to different characteristics of the wastewater and the different degrees of mixing. The field-scale ponds were evaluated at the Williamson Creek Plant at a lower influent BOD than that at the Govalle Wastewater Treatment Plant where the laboratory-scale tests were carried out. Therefore, the minimum possible soluble effluent BOD can be achieved in a completely mixed aerated lagoon which is operated at a detention time of about two days.

The effluent concentration of total BOD is markedly affected by the effluent concentration of suspended solids. The relationship between the effluent suspended solids and the effluent total BOD is presented in Figure 4-5. These data were recorded for the aerated lagoons located at Williamson Creek Plant. The average ratio of effluent total BOD to effluent suspended solids is about 0.55. The extended aeration plant at the Williamson Creek Plant was operated as an aerated lagoon with a detention time of 2.6 days for a short period of time. The effluent suspended solids concentration of the aerated lagoons and the final clarifier were less than 25 mg/l and 32 mg/l, respectively. The BOD concentration in the clarifier effluent was about ten mg/l. The total BOD in the effluent from the aerated lagoon operated at detention times of three to eight days ranged between 20 and 70 mg/l with an average value of 40 mg/l. However, at a one-day detention time the average total effluent BOD varied from 50 to 120 mg/l. It should be pointed out at this time that the power level and therefore the mixing level in the aerated lagoon was very high and almost no settling of any of the solids was observed.

The quality of the effluent of a lagoon that is not completely mixed and in which a portion of the suspended solids are permitted to settle depends on the power level, the detention time, the distance of the aerators from the effluent, as well as the type, size and construction of the effluent weirs. In some of these incompletely mixed tanks the concentration of algae which develop in the basin can also be a significant factor which must be considered in the design. As the power level decreases, the concentration of mixed liquor suspended solids must also decrease and the concentration of algae will tend to increase. If the effluent of aerated lagoons is stored in a quiescent pond to reduce the concentration of suspended solids, problems may develop with the growth of algae in the ponds resulting in suspended solids and possibly additional BOD in the effluent (McKinney and Benjes, 1965). Algae have been reported to grow well in aerated lagoons which have detention times as long as 20 to 30 days (Penman, 1970).

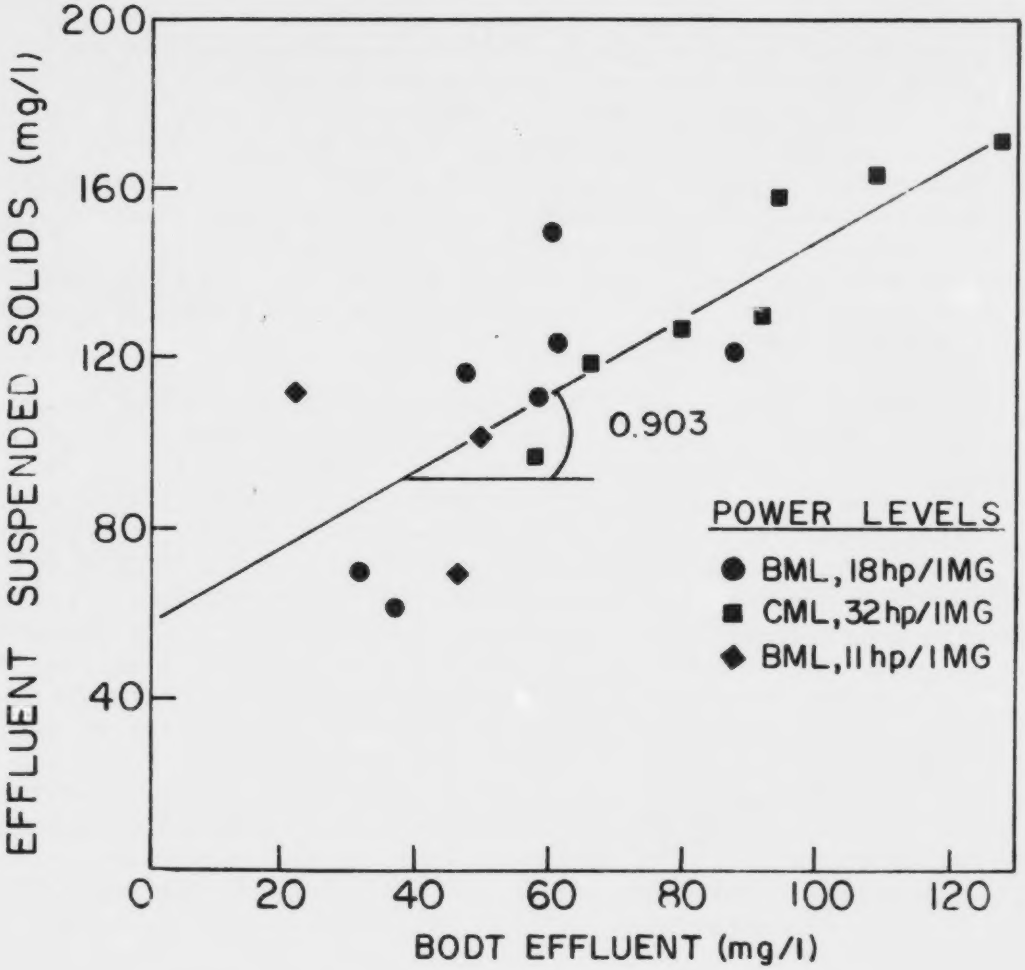


FIG.4-5

EFFLUENT SUSPENDED SOLIDS VERSUS EFFLUENT BODT

Published data which describe the performance of aerated lagoons treating municipal wastewater are relatively sparse. McKinney and Benjes (1965) reported that the effluent of an aerated lagoon treating municipal wastewater contained 50 mg/l of BOD and 130 mg/l of suspended solids. The average influent BOD concentration was about 230 mg/l and the detention time in the basin ranged from three to nine days. The BOD and suspended solids concentrations in the effluent of a series of three ponds which followed the aerated lagoon were 31, 20, and 14 mg/l for BOD and 55, 39, and 36 mg/l for suspended solids. Other data reported for a system including an aerated lagoon operating at detention times of 0.8 to 4.4 days followed by a stabilization pond indicate that the effluent from the aerated lagoons contained 58 mg/l of BOD while the effluent BOD from the stabilization pond was 28 mg/l. The influent BOD was 170 mg/l.

Penman (1970) reported that the effluent of aerated lagoons operated at detention times of 20 to 30 days contained an average BOD of 36 mg/l and and average suspended solids concentration of 36 mg/l. The influent wastewater to these ponds contained 175 mg/l BOD and 180 mg/l of suspended solids. These results indicate that a system including an aerated lagoon with a short detention time followed by a pond to remove the suspended solids can produce an effluent which is of equal quality or better to the effluent from aerated lagoons which are operated at long detention times.

Temperature Effects

The effects of temperature on the aerated lagoon process are more pronounced than on the activated sludge process, since the concentration of mixed liquor suspended solids in the aerated lagoon system is much lower than that maintained in an activated sludge process. The effect of temperature on the rate constant can be written as Equation 4-9.

$$k_{20} = k_T \theta^{(20-T)} \qquad 4\text{-}9$$

The coefficient θ will vary from 1.035 (Sawyer, 1968) to between 1.06 to 1.09 (Eckenfelder, 1969) and up to 1.1 according to Bartsch (1970). A value of 1.08 is generally accepted for aerated lagoons. Equation 4-1 can then be rewritten to include temperature changes as Equation 4-10.

$$S = \frac{1}{atk_T\, 1.08^{(20-T)}} + \frac{b}{ak} \qquad 4\text{-}10$$

The term b/ak does not change with temperature because both the endogenous respiration rate, b, and the rate constant, k, are temperature dependent and both terms are affected in the same way by changes in temperature. The optimum detention time for design purposes would be two days for an aerated lagoon operating at a temperature of 20°C (68°F). Longer detention times would be required at lower temperatures since the rate constant k_T decreases with decreasing temperature. The detention time required for operation at temperatures less than 20°C can be calculated from Equation 4-11.

$$t_T = t_{20}\, 1.08^{(20-T)} \qquad 4\text{-}11$$

It is interesting to note that the curves in Figure 4-4 indicate that temperature has very little influence on the effluent BOD at detention times longer than 2.5 days. However, the scatter of the data around the detention time of one day for the field lagoons is partially caused by temperature variations.

The temperature of the liquid in an aerated lagoon is generally different than the temperature of the influent wastewater. Bishop, Malina, and Eckenfelder (1971) derived the following mathematical model for predicting temperatures of the lagoon, and this model is presented as Equation 4-12.

$$T_L = \frac{8.34\, QT_i + 145A\,(T_A - 2)}{145A + 8.34\, Q} \qquad 4\text{-}12$$

in which:

T_L = weekly average lagoon temperature (°F)

T_i = weekly average influent temperature (°F)

T_A = weekly average air temperature (°F)

Q = wastewater flow (gal/day)

A = surface area of the lagoon (sq ft)

145 = average heat exchange coefficient (BTU/sq ft - day - °F)

This model can be converted to a dimensionaless form expressed in Equation 4-13.

$$\frac{T_L}{T_i} = \frac{8.34\,(\frac{Q}{A}) + 145\,(\frac{T_A - 2}{T_i})}{145 + 8.34\,(\frac{Q}{A})} \qquad 4\text{-}13$$

A nomograph has been developed relating the ratio of the air temperature to the influent temperature $[(T_A - 2)/T_i]$ as well as the hydraulic surface loading (Q/A) expressed as (gal/sq ft/day). The relationship of detention time is also included in Figure 4-6 for a lagoon depth of ten feet. The curves indicate that at a short detention time the temperature has a relatively small influence. Bishop, Malina, and Eckenfelder (1971) have shown that this model was relatively insensitive with respect to the heat exchange coefficient. The term $(T_A - 2)$ approximates the equilibrium temperature which is a temperature that a mixed water body without influent or effluent reaches when exposed to air.

Oxygen Uptake

The oxygen uptake for a completely mixed aerated lagoon will follow the basic relationship expressed as

$$\frac{R}{X} = a' \frac{S_o - S}{Xt} + b' \tag{4-14}$$

The value of S in Equation 4-14 is the nonbiodegraded fraction of the influent, and it is not the measured effluent BOD concentration. It can be assumed that at a theoretical detention time longer than one day about 90 to 95 percent of the initial BOD has been degraded biologically and that $a' = 0.53$ lb/lb. Therefore, $0.53\ (S_o - 0.05\ S_o) = 0.53\ (0.95)\ S_o = 0.5\ S_o$ and Equation 4-14 can be rewritten in a similar manner to that for the activated sludge process as Equation 4-15.

$$\frac{R}{X} = 0.5 \frac{S_o}{Xt} + b' \tag{4-15}$$

Since the active mixed liquor suspended solids concentration is not readily measurable in the aerated lagoon system, Equation 4-15 can be simplified.

$$R = 0.5 \frac{S_o}{t} + b'X \tag{4-16}$$

However, the active mixed liquor suspended solids concentration can be assumed to be approximately equal to $0.5\ S_o$. This value is a function of the power level and for a completely mixed aerated lagoon Equation 4-16 can be rewritten.

$$R = 0.5\ S_o \left(\frac{1}{t} + b\right) \frac{8.34}{1000} \tag{4-17}$$

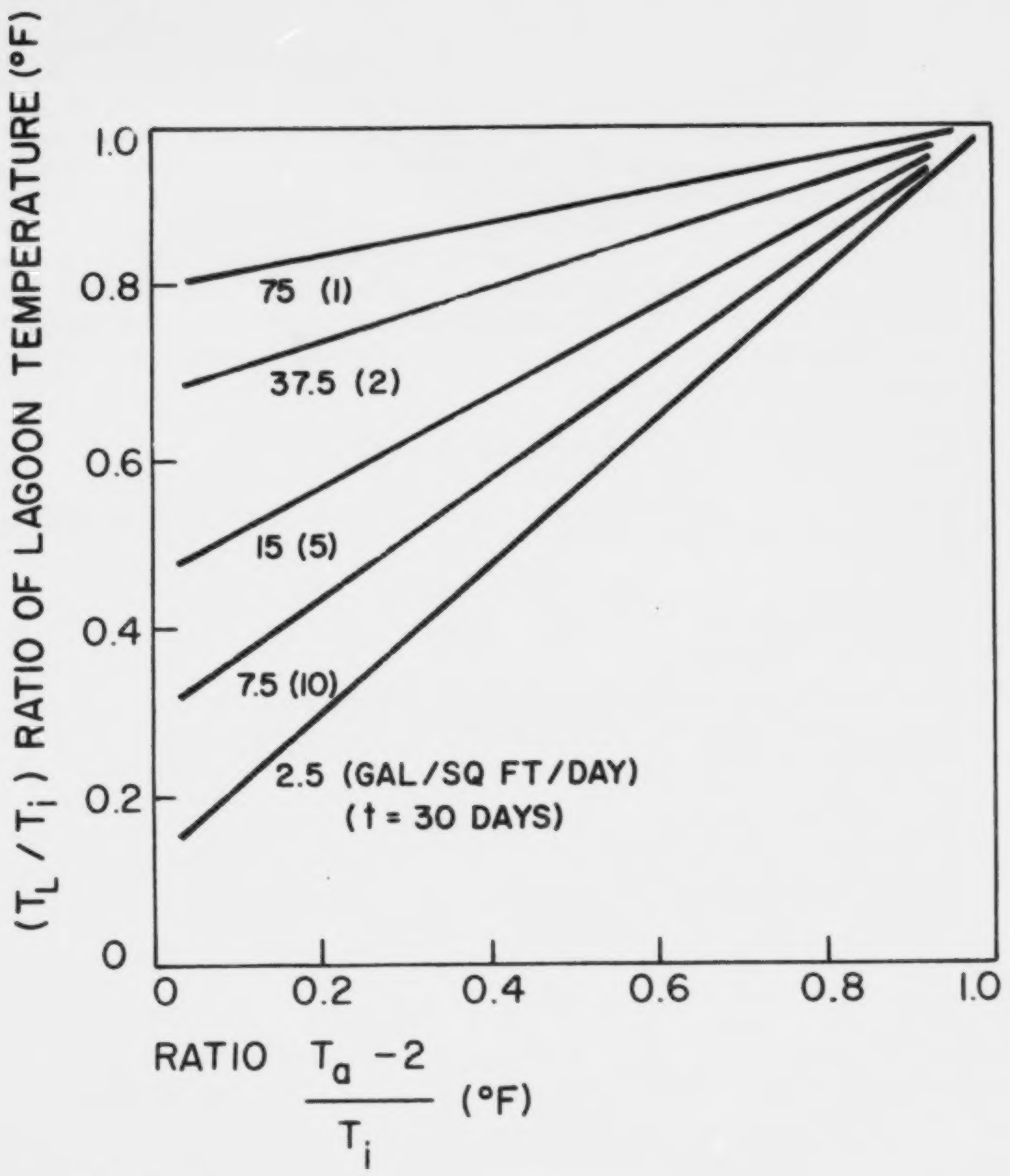


FIG. 4-6
AERATED LAGOONS
TEMPERATURE PREDICTION NOMOGRAPH

in which:

R = oxygen uptake per unit volume (lb/1000 gal - day)

S_o = total BOD concentration in the influent (mg/l)

b' = endogenous uptake rate (day^{-1}) ($b' = 0.15$ for municipal wastewaters)

In aerated lagoons in which sludge deposits undergo anaerobic decomposition, the soluble organic compounds released during anaerobic decomposition will be introduced into the bulk liquid and exert an additional oxygen demand. The rate of anaerobic decomposition is higher at the higher temperatures during the summer months and relatively low during the winter months. Eckenfelder (1970) introduced factors of 1.05 and 1.2, to account for the feedback of oxygen demanding substances from the anaerobic decomposition of the settled sludge for winter and summer conditions, respectively. Therefore, Equation 4-17 can be rewritten for summer conditions and winter conditions, respectively.

$$R = 0.5\, S_o \left(\frac{1}{t} + b\right) \frac{8.34}{1000} (1.2) \text{ (Summer)} \qquad 4\text{-}18$$

$$R = 0.5\, S_o \left(\frac{1}{t} + b\right) \frac{8.34}{1000} (1.05) \text{ (Winter)} \qquad 4\text{-}19$$

Mixing and Power Levels

The degree of mixing has the greatest influence on the overall performance of aerated lagoons. At high mixing levels all of the suspended solids are maintained in suspension; therefore, the effluent suspended solids and total BOD concentrations are relatively high. At the lower mixing level, a portion of the solids settle to the bottom of the lagoon, and the concentration of suspended solids in the effluent is much lower than that observed at the higher mixing levels. Therefore, the effluent concentration of total BOD is also considerably lower. The degree of mixing in an aerated lagoon is a function of a type of aeration equipment employed and the power level expressed as HP/1000 gal.

The relationship between the concentration of suspended solids in the aerated lagoon and the power level was evaluated at the aerated lagoons at the Williamson Creek Wastewater Treatment Plant in Austin, Texas. The aerated lagoon was operated at the highest possible power level for a period of time. After the suspended solids concentration in a lagoon was determined the power level was reduced in the evening. The suspended solids concentration in the lagoon were determined at two different times of the next day at several sampling points. This procedure was repeated until the lowest power level was applied.

The results of this investigation are presented in Figure 4-7 which relates the concentration of mixed liquor suspended solids to power level. The data indicate that there is a sharp decrease in the suspended solids concentration as the power level decreases. However, at a minimum power level the concentration of a mixed liquor suspended solids reached an equilibrium value. The concentration of mixed liquor suspended solids was about 55 mg/l at power levels, between 9 and 13 horsepower per million gallons. The data also indicate that the highest possible power level for the aerator and pond geometry was 20 horsepower per million gallons. At this power level all the solids in the system were not in suspension, since theoretically, the maximum mixed liquor suspended solids concentration would have been approached asymptotically. However, in this particular aerated lagoon system, the maximum mixed liquor suspended solids concentration could have been only slightly higher than the value of 130 mg/l indicated in Figure 4-7 since the influent suspended solids concentration was not significantly greater in this value.

The mixed liquor suspended solids concentration under steady state operating conditions for an aerated lagoon is affected by the influent suspended solids, the detention time and the degree of mixing. The data observed in field-scale ponds and presented in Figure 4-8 indicate that a wide range of possible effluent suspended solids concentrations can be observed at a given power level. There is a slight increase in the effluent suspended solids concentration as the power level increases; however, this increase is not markedly obvious. A comparision of the suspended solids concentration in the aerated lagoon as a function of the power level with results of mixing studies reported by Bishop, Malina, and Eckenfelder (1971) provides an interesting insight into the problem of mixing aerated lagoons. The degree of mixing is expressed as a number of completely mixed tanks in a series as discussed earlier. A tank is completely mixed as the calculated number of tanks in series is one. An infinite number of tanks in series represents a plug flow system. The relationship between the number of completely mixed tanks in series and the power level for the aerated lagoons at the Williamson Creek Treatment Plant is presented in Figure 4-9. The data indicate that the number of completely mixed tanks in series is approximately one at power levels between 30 and 50 horsepower per million gallons. As the power level decreases below 30 horsepower per million gallons, the number of completely mixed tanks in series increases and at power levels below 20 horsepower per million gallons the number of completely mixed tanks in series increases sharply. The results of tracer studies as well as the relationship between mixed liquor

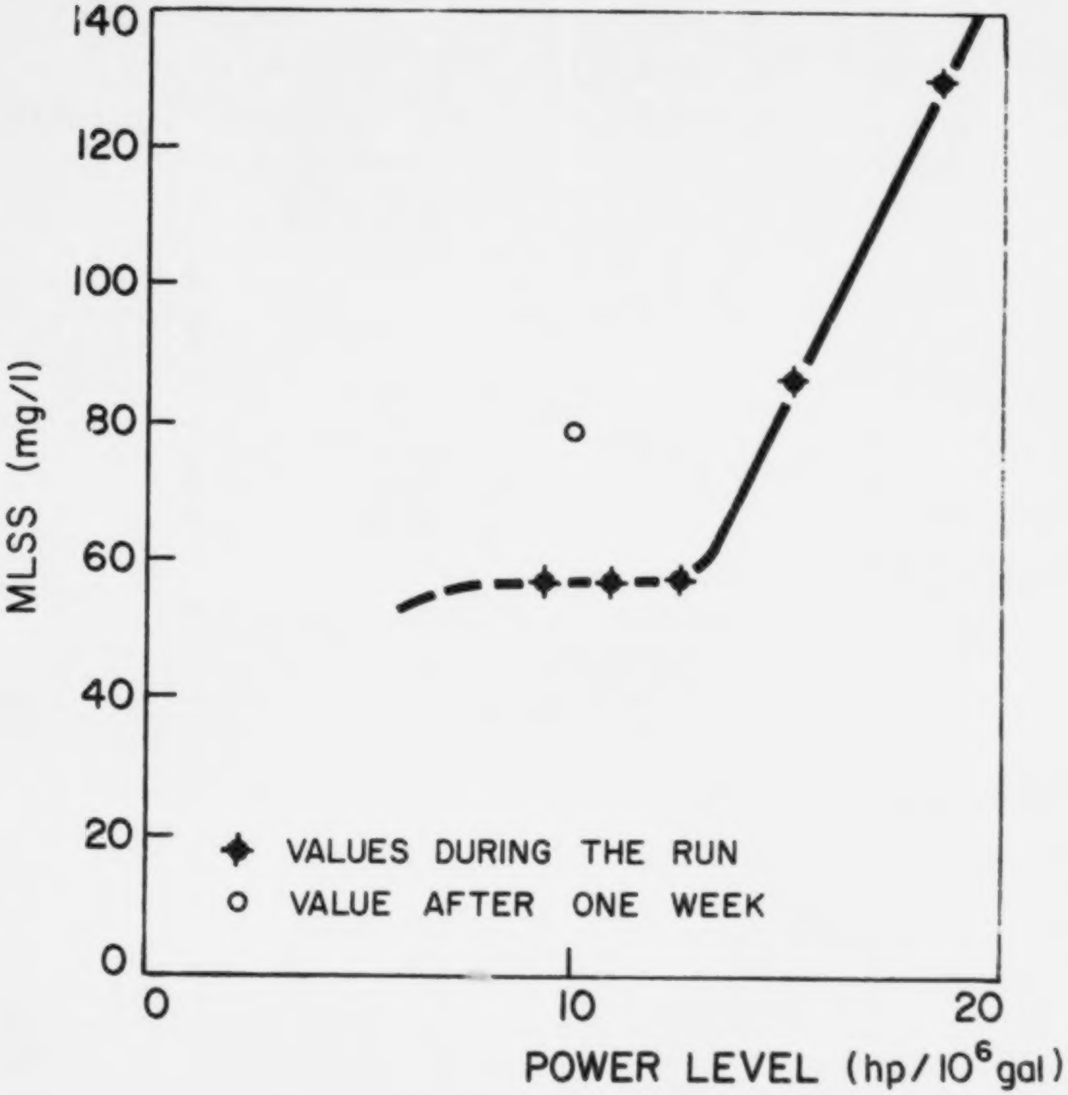


FIG.4-7

POWER LEVEL VERSUS MLSS

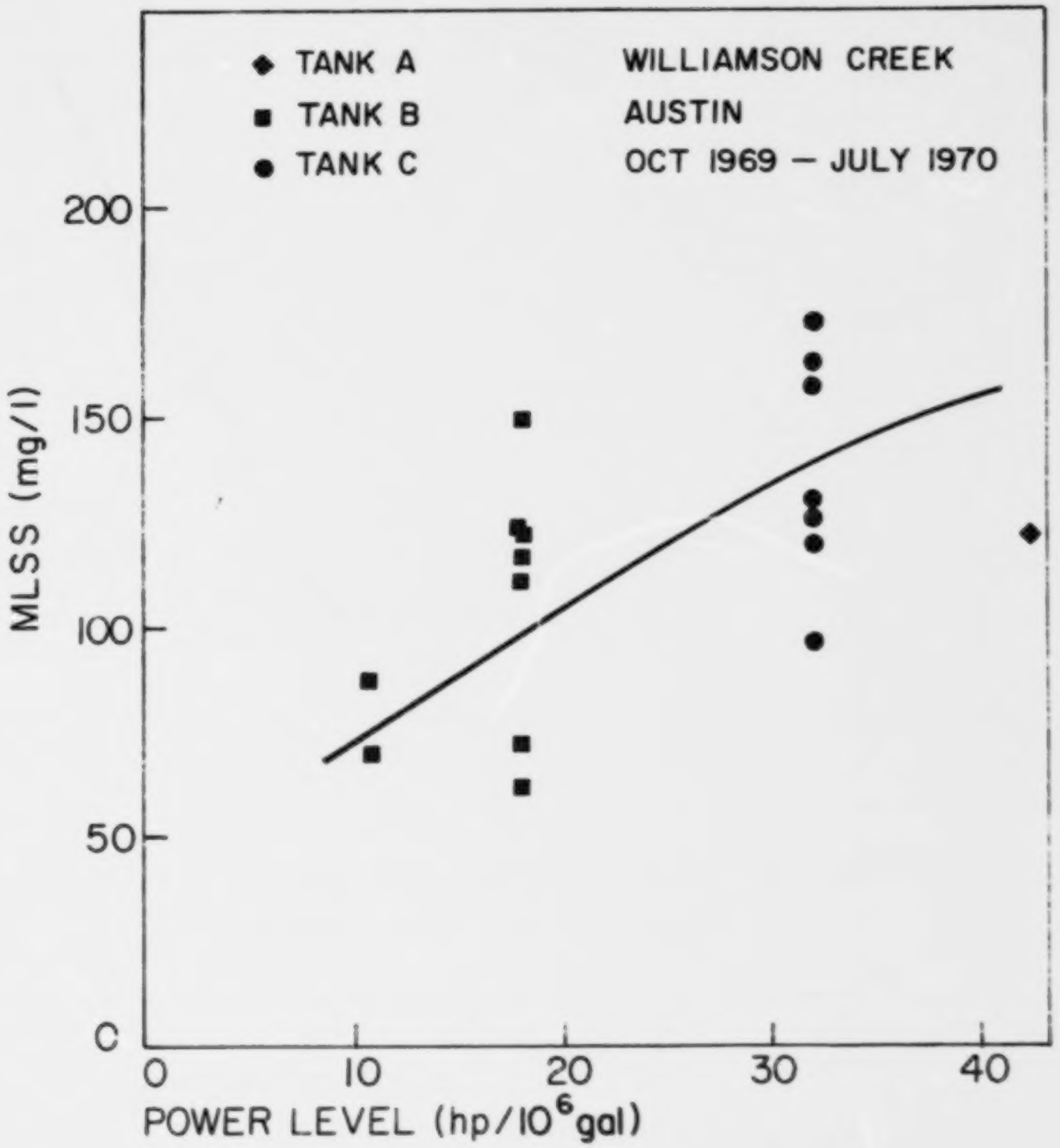


FIG. 4-8

MLSS AND POWER LEVEL

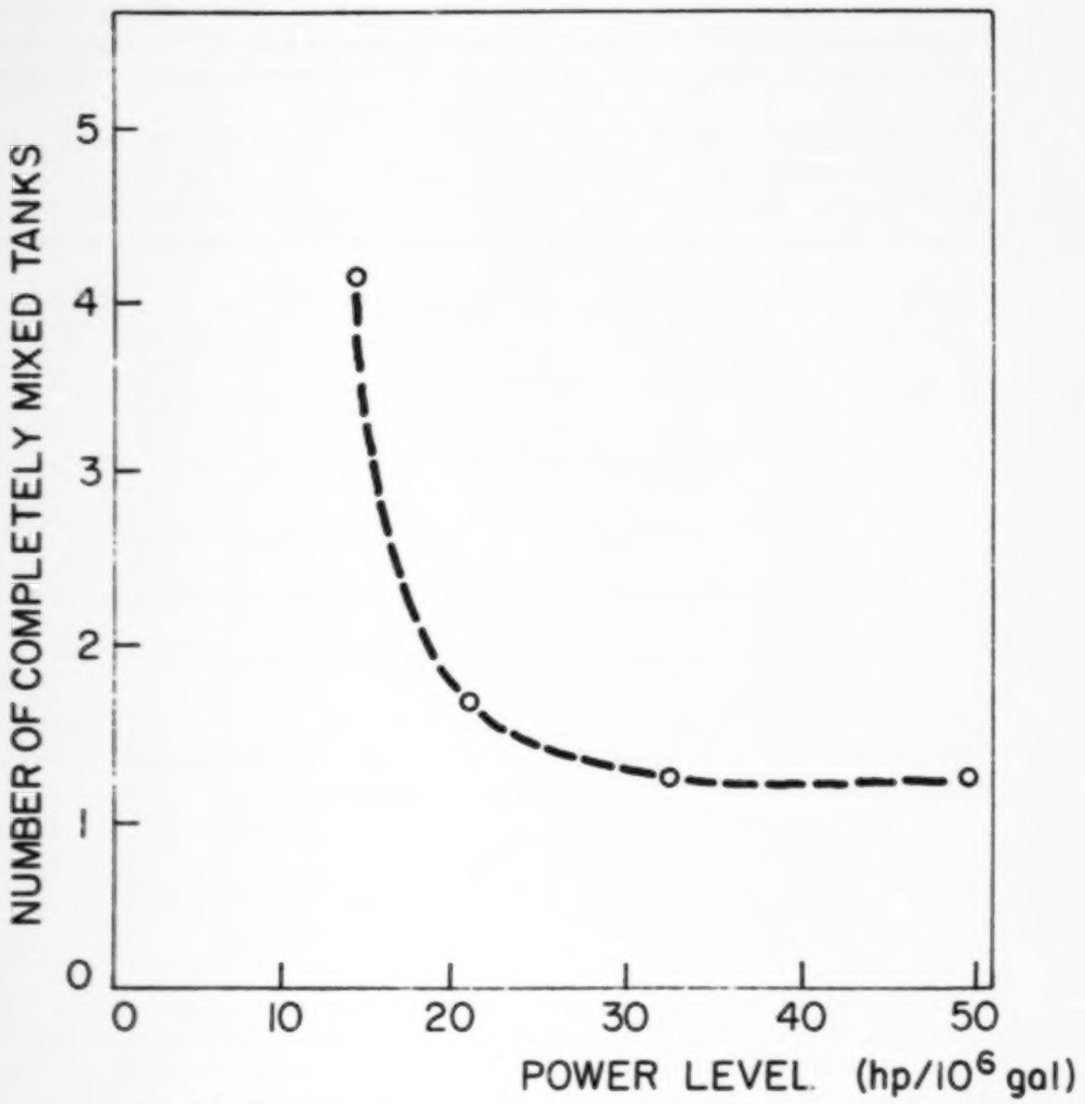


FIG. 4-9

POWER LEVEL AND DEGREE OF MIXING

suspended solids concentration and power level substantiate the assumption by Eckenfelder (1970) that aerated lagoons can be considered completely mixed tanks at power levels greater than 30 horsepower per million gallons.

The power level required to provide sufficient oxygen transfer to satisfy the oxygen uptake rates can be calculated based on the influent BOD and the detention time using Equation 4-17. The oxygen transfer rate must be equal to the oxygen uptake rate and may be calculated using Equation 4-20.

$$OT = \frac{C_s - C}{C_{s_{20}}} (N_o) (P_v) (24) (1.02^{(20-T)}) \qquad 4\text{-}20$$

in which:

OT = oxygen transfer rate (lb/1000 gal-day)

C_s = concentration of dissolved oxygen at saturation (mg/l)

C = concentration of dissolved oxygen in the lagoon (mg/l)

N_o = oxygen transfer efficiency (lb/HP-hr)

Combining Equations 4-17 and 4-20 which are equal and solving for the power level results in Equation 4-21 which can be used for different detention times and for different influent BOD loadings.

$$P_v = 1.73 \times 10^{-4} \left(\frac{C_{s_{20}}}{C_s - C}\right) \left(\frac{S_o}{N_o}\right) \left(\frac{1}{t} + 0.15\right) (1.02)^{(T-20)} \qquad 4\text{-}21$$

The curves presented in Figure 4-10 are based on an oxygen transfer efficiency $N_o = 1.7$ lb of O_2 per hp-hr, a dissolved oxygen concentration under summer conditions of 7.0 mg/l and a requirement of a minimum dissolved oxygen concentration in a lagoon of 2.0 mg/l. The power level calculated using Equation 4-21 is that required to maintain oxygen transfer at a rate sufficient to satisfy the oxygen uptake requirements. The series of curves presented in Figure 4-10 indicate that the power level required to satisfy the oxygen uptake requirements in a lagoon operating at a one day detention time and treating municipal wastewater with BOD concentrations between 100 and 150 mg/l is approximately 35 to 40 horsepower per million gallons. This power level is also sufficient to maintain all suspended solids in suspension. Therefore, the lagoon is completely mixed and aerobic. However, if the detention time in a lagoon treating the same wastewater is increased to two days, the power level required to provide

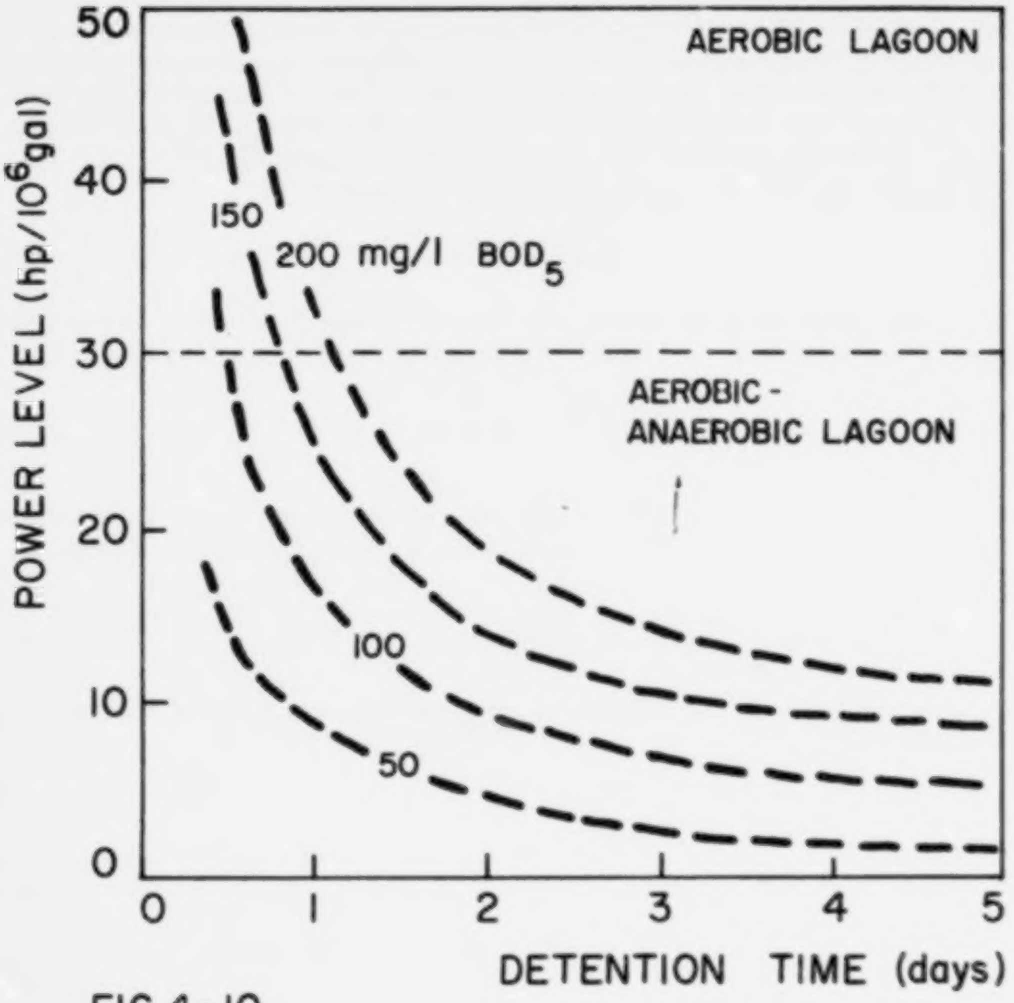


FIG. 4-10

POWER LEVEL FOR OXYGEN TRANSFER

oxygen to meet the oxygen uptake requirement is reduced to between 10 and 15 HP/million gallons. At this lower power level in the larger lagoon the intensity of mixing would not be sufficient to completely mix the basin and settling of the suspended solids will occur.

The results of the mixing studies indicate that a certain degree of bypassing occurred in the lagoons at the Williamson Creek Plant. Relatively high concentrations of tracer were measured in the effluent consistently at about 10 to 20 minutes after the tracer input. Therefore, to minimize the short-circuiting or bypassing in a square tank in which a single surface aerator is located in the center of the tank, it would be advantageous to introduce the influent wastewater directly below the aerators.

Effluent Suspended Solids

The suspended solids concentration in the effluent of a completely mixed aerated lagoon can be calculated using a modification of Equation 2-22.

$$\frac{1}{G} = 0.6 \frac{(S_o + X_o)}{Xt} - 0.075 \qquad 4\text{-}22$$

However, in the aerated lagoon process the sludge age (G) is approximately equal to the detention time in the basin; therefore, Equation 4-22 can be modified

$$X = 0.6 (S_o + X_o) - 0.075\, Xt \qquad 4\text{-}23$$

The concentration of mixed liquor suspended solids can be calculated from Equation 4-24.

$$X = \frac{0.6 (S_o + X_o)}{1 + 0.075\, t} \qquad 4\text{-}24$$

in which:

S_o = total influent BOD concentration (mg/l)

X_o = influent suspended solids concentration (mg/l)

t = theoretical detention time (days)

X = effluent suspended solids concentration (mg/l)

The results of laboratory-scale experiments in which completely mixed aerated lagoons were operated verify the applicability of Equation 4-24 to predict the effluent suspended solids concentration.

The effluent suspended solids concentration in the aerated lagoons in which a portion of the suspended solids settle is a function of the power level, the location of the effluent, the construction of the effluent structure (baffles, weirs, stilling basins) and the detention time. A rough estimate of the effluent suspended solids concentration for this type of lagoon can be obtained from the curve in Figure 4-7 which relates the effluent suspended solids concentration to the power level.

DESIGN FACTORS

The basic data required for the design and layout of an aerated lagoon system to treat municipal wastewater include:

(a) wastewater flow (MGD)
(b) average influent BOD concentration (mg/l)
(c) average influent suspended solids concentration (mg/l)
(d) the temperature of the wastewater including the lowest and highest average weekly liquid and air temperatures.

Experimental results and published data indicate that a high reduction in the concentration of soluble BOD can be achieved in aerated lagoons operating at a relatively short detention time of about two days and at a temperature of about 20°C. The effluent total BOD and suspended solids of aerated lagoons operated at detention times of one to eight days are generally high when compared with the usual effluent requirements. The effluent suspended solids generally settle well. An aerated lagoon with a three day detention time followed by a clarifier will result in a final total BOD of 15 mg/l or less. However, the total BOD of the aerated lagoon effluent which contains the biodegradable settleable suspended solids will be considerably higher than the 15 mg/l. Aerated lagoons operated at detention time in excess of 20 to 30 days, are very similar to waste stabilization ponds.

Proper design and location of the aeration equipment will enable the system to produce an acceptable effluent. However, land requirements for this type of system are greater than for the activated sludge system. Aerated lagoons can effectively treat municipal wastewater in relatively short detention times; however, a portion of the effluent suspended solids must be removed to reduce the total BOD to the acceptable levels. An aerated lagoon can be followed by a waste stabilization pond, by another aerated lagoon operated at a very low power level, or by using a final clarifier from which the concentrated sludge may be pumped to a disposal site. A schematic representation of these various alternatives is presented in Figure 4-11.

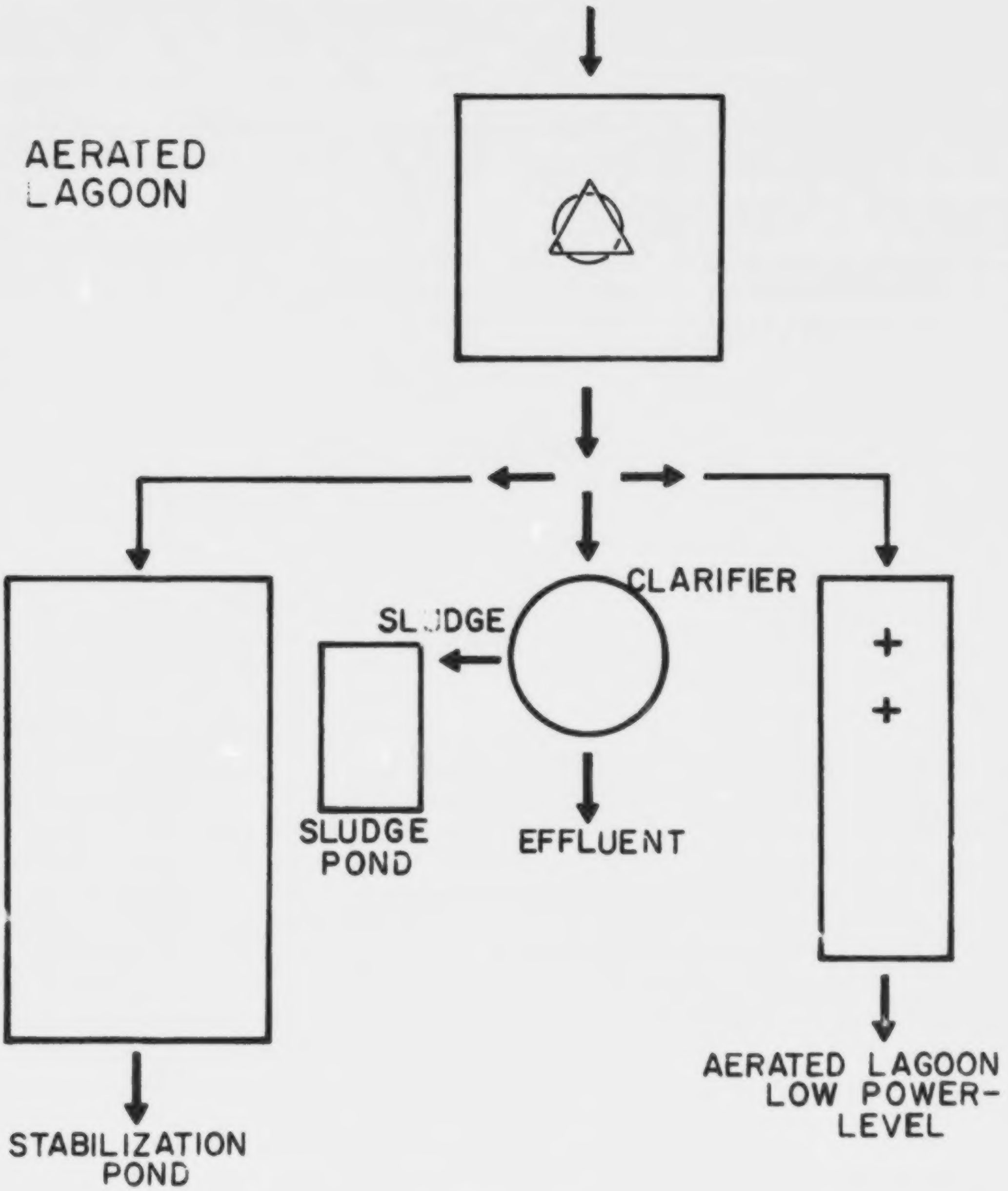

AERATED
LAGOON
CLARIFIER
SLUDGE
SLUDGE
POND
EFFLUENT
AERATED LAGOON
LOW POWER-
LEVEL
STABILIZATION
POND

FIG. 4-11

AERATED LAGOON SYSTEMS

DESIGN PROCEDURE

The following procedure for the design of an aerated lagoon system should result in effective treatment of municipal wastewater:

(a) Select a detention time which will result in a low residual soluble BOD. A detention time of two days at an operating temperature of 20°C should be satisfactory for the treatment of municipal wastewater at an initial total BOD of 100 to 150 mg/l.

(b) The selected detention time should be modified for temperature changes. The minimum lagoon temperature can be computed from the curves in Figure 4-6 using the minimum expected air and wastewater temperatures. The required detention time can be determined based on Equation 4-11. This approach requires a trial and error procedure.

(c) The volume of pond required can then be calculated ($V = Qt$).

(d) The power level required for oxygen transfer to satisfy the oxygen uptake requirements can be calculated from Equation 4-21. It should be pointed out that this calculated power level in most cases only is sufficient to satisfy the oxygen requirements for biodegradation.

(e) Compare the power level required for oxygen transfer with that required to maintain an effluent suspended solids concentration (mg/l) about equal to the influent BOD concentration; e.g. if the influent BOD is equal to 100 mg/l, the effluent suspended solids concentration can also be about 100 mg/l. The curve in Figure 4-7 indicates that a power level of about 20 horsepower per million gallons would be sufficient to maintain an effluent suspended solids concentration of about 100 mg/l. If a completely mixed aerobic lagoon is desired, the required power level for mixing should be equal to or more than 30 horsepower per million gallons.

(f) In the case of the aerated lagoon in which some settling will take place, the design of the aerators should be based on the higher of the two required power levels. The total power requirement for the aerators can be calculated based on the total volume of the aerated lagoon.

(g) Select the sludge handling system based on the method of removal of suspended solids from the effluent.

TRICKLING FILTER PROCESS

PROCESS DESCRIPTION

A trickling filter consists of a bed of coarse material such as broken stones, clinkers, wood slats, plastic tubes, corrugated plastic sections, or other material over which wastewater is distributed. Wastewater flows over the medium on which a zoogleal slime develops. Dissolved organic material in the wastewater is transported into the slime where biological oxidation takes place and the effluent liquid is collected in an underdrain system. Air passes through the void spaces in the medium and supplies the oxygen required to maintain an aerobic environment. A trickling filter will operate properly as long as the void spaces are not clogged by the influent solids or by excessive slime growth. In a well operating filter the wastewater flows in a vertical direction with very little crossflow. The wastewater is applied in such a way that the zoogleal mass is alternately in contact with the wastewater and air.

Trickling filters have been operated over a wide range of hydraulic and organic loadings. The classical definitions of a high-rate and a low-rate filter are presented in the Manual for Sewage Treatment Plant Design (1967). However, the classical definitions are not related to the performance of the trickling filter; therefore, these relationships are not very applicable to the design of trickling filter plants.

The need for primary sedimentation prior to the trickling filter is a function of the concentration of suspended solids in the influent wastewater as well as the effective size of the medium. Stone medium generally requires effective primary clarification to minimize problems with clogging. Primary sedimentation is not necessary for those systems in which the medium consists of corrugated plastic or other media which have large voids.

Various alternate flow diagrams for trickling filter plants are illustrated in Figure 5-1. Trickling filters may be operated with recycling of part of the effluent as well as by having a number of filters in series. The suspended solids in the effluent of the trickling filter are biological materials which slough off the medium. This sloughing action takes place as the slime grows to such an extent that the hydraulic shear force resulting from the downward flow of the wastewater actually separates the slime from the surface of the medium and carries it to the underdrain system.

Trickling filters are relatively simple to operate, since the primary control of the performance involves the rate of recirculation of filter effluent.

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Trickling filters have been reported to recover readily from shock loads, however, the performance of trickling filters is affected much more noticeably by shock loadings than that of the activated sludge process. The applicability of the trickling filter process may be summarized by the following statement from the Manual of Design of Sewage Treatment Plants (1967): "They are capable of providing adequate treatment of such wastes where the production of a plant effluent of 20 to 30 milligrams per liter of BOD is acceptable or where partial treatment is acceptable."

PERFORMANCE OF TRICKLING FILTERS

The flow pattern in a trickling filter is generally assumed to be plug flow. However, the results of tracer studies indicate that some mixing takes place, but no backmixing is possible. Some particles of liquid will pass through the filter at a much more rapid rate than others. In spite of this mixing the mechanism of biodegradation of the soluble BOD can be described by first order reaction kinetics applied to a plug flow process.

The overall performance of a trickling filter involves a number of processes including coagulation, flocculation, and/or biodegradation of the particulate organic matter as well as the efficiency of the final clarifier in separating the suspended solids from the liquid. Eckenfelder (1961) correlated trickling filter data using a retardant reaction which is a first order reaction applied to a completely mixed system.

The basic mathematical model describing the biodegradation of soluble substrate in a plug flow system was presented as Equation 2-13, which is rewritten as Equation 5-1:

$$\frac{S_e^*}{S_o^*} = e^{-kXt} \qquad 5\text{-}1$$

in which

S_e^* = effluent concentration of soluble substrate (mg/l)

S_o^* = influent concentration of soluble substrate (mg/l)

k = rate constant (time^{-1})

X = concentration of active microorganisms (mg/l)

t = time of flow through the filter

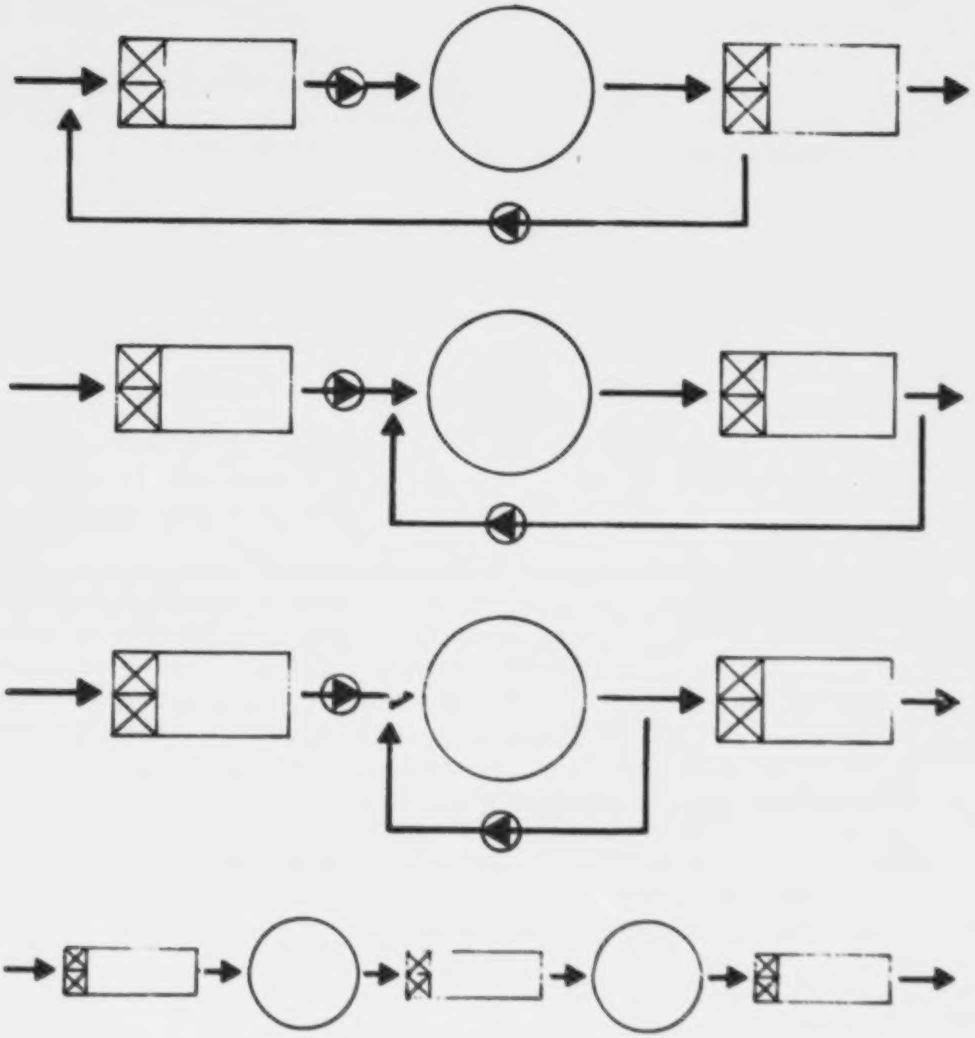

FIG. 5-1

TRICKLING FILTER; FLOW DIAGRAMS

If it is assumed that the entire filter medium is covered with a uniform layer of microorganisms, the total microbial population is a function of the available surface and can be expressed as Equation 5-2.

$$X = c_1 A_v \tag{5-2}$$

in which

X = active microbial population per unit volume of filter

A_v = specific surface of medium (sq ft/cu ft)

c_1 = constant

The factor c_1 is a function of the thickness of the film, the density of the microbial population, the penetration of oxygen, and the transport of substrate. Oxygen and substrate are transported into the film by means of diffusion processes. The aerobic surface of a thin film is normally underlain by an anaerobic layer. Therefore, a very thick film does not necessarily indicate that the number of active aerobic microorganisms is very large. The soluble organic compounds released during anaerobic degradation are transported into the aerobic film.

The flow-through time is a function of the real flow per unit area of nominal surface area of the filter medium and may be calculated from Equation 5-3, if the flow velocity is assumed to be independent of depth.

$$t = c_2 DQ^{-n} \tag{5-3}$$

in which

t = time of flow to travel the depth of the filter

Q = hydraulic surface loading (million gallons per acre per day (MGAD))

D = filter depth (feet)

c_2, n = constants which characterize the medium

The constants n, and c_2, are a function of the size and configuration of the medium. The value of n is relatively constant for several filter media; however, the coefficient c_2 is a function of the specific surface areas

$(c_2 = c_3 A_v^m)$. In practice it is very difficult to independently evaluate the exponents m and n. Equation 5-3 can be rewritten as Equation 5-4.

$$t = c_3 D A_v^m Q^{-n} \qquad 5\text{-}4$$

The biodegradation rate constant, k, the coefficient, c_1, and the specific surface area are interdependent and cannot be effectively evaluated separately. The coefficient, c_2, which relates the time of flow through the filter with the depth and the hydraulic loading can be determined from tracer studies using tracer response techniques. However, when these coefficients are determined from biological experimentation, a single constant is developed which can be described by Equation 5-5.

$$k^* = k\, c_1\, c_2 \qquad 5\text{-}5$$

Equation 5-1 can be modified and used to calculate the effluent substrate concentration.

$$S_e^* = S_o^* e^{-(k^* A_v\ DQ^{-n})} \qquad 5\text{-}6$$

This first order reaction is widely used to describe the performance of trickling filters. The substrate removal rate constant, k* is a function of the configuration of the filter medium, the thickness of the active film, and the characteristics of the influent wastewater as well as temperature. Therefore, this rate constant is different for various types of wastes and types of filter medium. The exponent, n, has been reported to range in value from 0.30 to 0.90. The numerical value of this exponent decreases as the slime thickness increases, and the flow-through time increases as the slime thickness increases. Therefore, it is essential that the exponent, n, is evaluated under process conditions in which a slime layer has been developed rather than using tap water and clean filter medium.

Eckenfelder (1966) reported a method for the determination of a rate constant, k*, and the exponent, n. The exponent in Equation 5-6 can be simplified and used for the evaluation of experimental data. At a constant flow rate of the same wastewater, for a specific filter medium, and at a constant temperature, the exponent can be reduced to

$$k^* A_v Q^{-n} = 2.303 k_Q \qquad 5\text{-}7$$

Equation 5-6 can be rewritten in logarithmic form and evaluated graphically.

$$\log_e (S_e^*/S_o^*) = 2.303 k_Q D \qquad 5\text{-}8$$

$$\log_{10}(S_e/S_o) = -k_Q D \qquad 5\text{-}9$$

The rate constant k_Q is the slope of the line resulting from a plot of the fraction of BOD remaining (S_e/S_o) at various depths of filter medium. This particular plot is shown in Figure 5-2. The data indicate that the rate constant, k_Q, is a function of the hydraulic loading since the slope of the line is different for different flow rates. The relationship between the rate constant and the hydraulic loading rate can be obtained by rearranging and rewriting Equation 5-7.

$$\log k_Q = \log\left(\frac{k^* A_v}{2.303}\right) - n \log Q \qquad 5\text{-}10$$

The exponent, n, is the slope of the line resulting from the plot of k_Q and the hydraulic loading, Q, shown in Figure 5-2. The rate constant of k^* can be calculated from Equation 5-10 since, at a hydraulic loading of one, log Q is zero and

$$\log k_Q = \log\left(\frac{k^* A_v}{2.303}\right) \qquad 5\text{-}11$$

Therefore, by extrapolating the line to a hydraulic loading of one the numerical value for k_Q can be graphically determined. The specific surface area is constant for the particular medium; therefore, the rate constant, k^*, can be determined from Equation 5-11 or graphically. The graphical method is shown by the third plot in Figure 5-2 in which the fraction of substrate remaining (S_e/S_o) is plotted versus the quantity ($A_v D Q^{-n}$). The slope of the line resulting from this plot is the rate constant, k^*.

The exponent, n, is dimensionless; however, the rate constant, k^*, is dependent upon the term Q^{-n} and is markedly affected by the units of the hydraulic loading. For example, a different numerical value will result if the hydraulic loading is expressed in terms of gallons per minute per square foot (gpm/sq ft) or million gallons per day per acre (MGAD) or cubic meters per square meter day (m^3/m^2 - day). The conversion from one set of units to the next is presented in Equation 5-12.

$$\text{MGAD} = 0.016\ \text{gpm/sq ft} = 0.939\ m^3/m^2 - \text{day} \qquad 5\text{-}12$$

The influence of the exponent, n, and the units in which the hydraulic loading is expressed are summarized in Table 5-1.

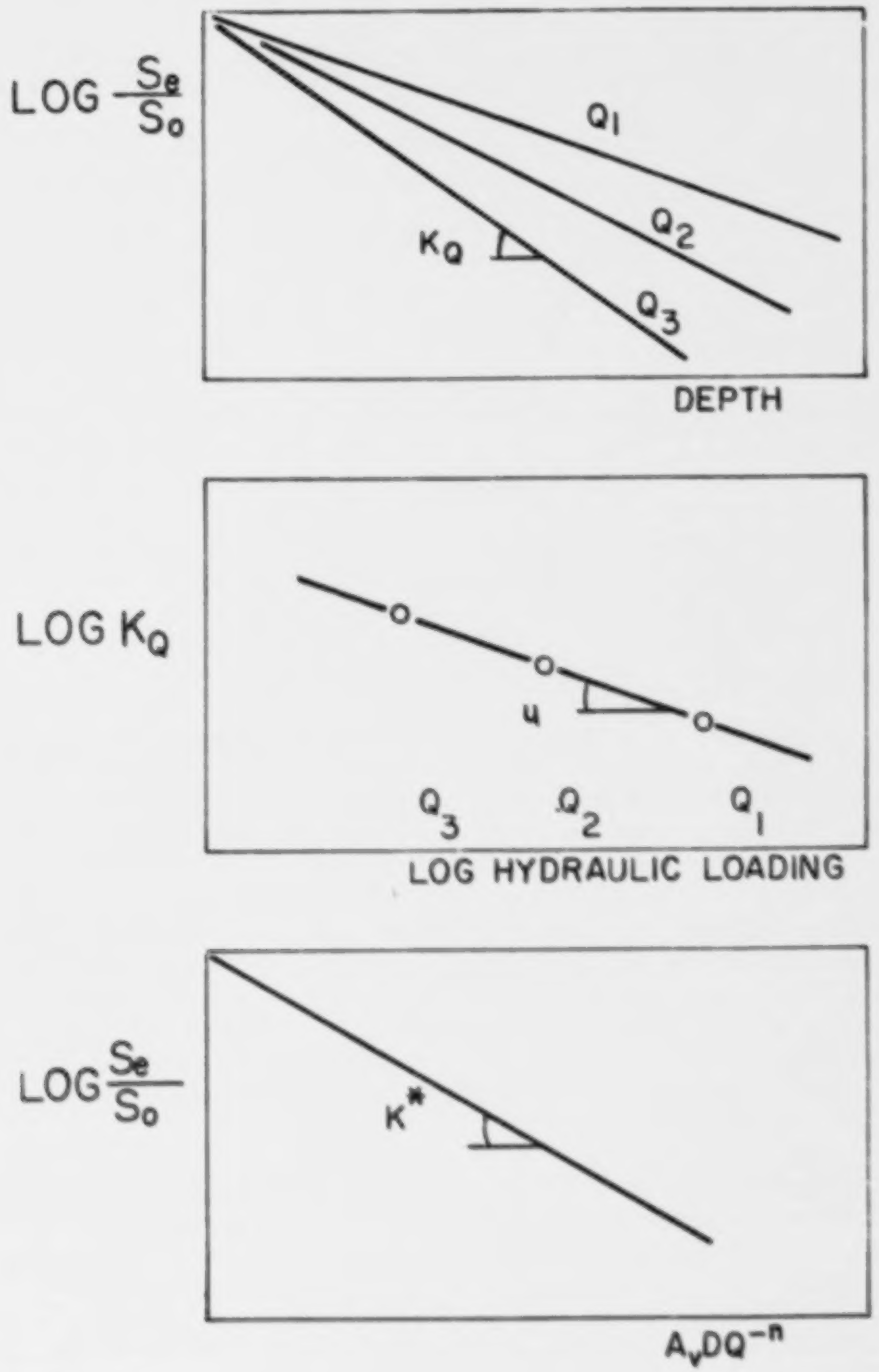

FIG.5-2
TRICKLING FILTER EVALUATION

Table 5-1 Influence of n and Q on Rate Constant k*

n/Q	MGAD	gpm/sq ft	m^3/m^2 - day
	k*	k*	k*
0.7	1.0	0.055	0.955
0.4	1.0	0.19	0.974

Additional confusion is introduced when values of the rate constant, k*, are presented without indicating the logarithmic base on which the values were determined. The rate constant, k*, derived from the slope of the plot is in units of base 10 logarithms; therefore, $k_e^* = 2.303\ k_{10}^*$. In the mathematical models describing the performance of trickling filters all equations are presented as exponential functions to the base e. Therefore, the rate constant, k*, should be expressed in terms of the base e along with the units of the hydraulic loading. The numerical value of the exponent, n, should also be included with the other data.

The influence of the exponent, n, on the rate constant, k*, is illustrated graphically in Figure 5-3. Experimental data at two different hydraulic loadings to a 20-foot deep filter are used to calculate the rate constant. The exponent, n, was evaluated from tracer studies and n = 0.70. Using this value of the exponent the hydraulic loading did not markedly affect the performance of the filter and the slope of the line through the data resulted in a single rate constant, k* = 0.056. The exponent reported in the literature is n = 0.40. Separate lines fit the data observed at each of the two different hydraulic loadings. These curves are also illustrated in Figure 5-3. The slopes of the lines were k* = 0.0121 and 0.0168, respectively when the exponent was n = 0.40. This illustration may provide a partial explanation for the relatively wide range of values reported in the literature for the rate constant k*.

Data observed using a three-foot diameter pilot-scale trickling filter with various media are presented in Figure 5-4. An exponent n = 0.50 was used in calculating the data for presentation in this figure. The specific surface area of the medium was 10 sq ft/cu ft of fresh limestone which had an effective size of four to six inches, 27 sq ft/cu ft of corrugated plastic sheets and 30 sq ft/cu ft of plastic rings. The data observed in the experiments in which the filter medium was either the corrugated plastic sheet or the plastic rings compare quite well. However, it is interesting to note that the slope of line drawn through the data will vary and depends on which data points might be selected. The composite sample data were used to draw the line of best

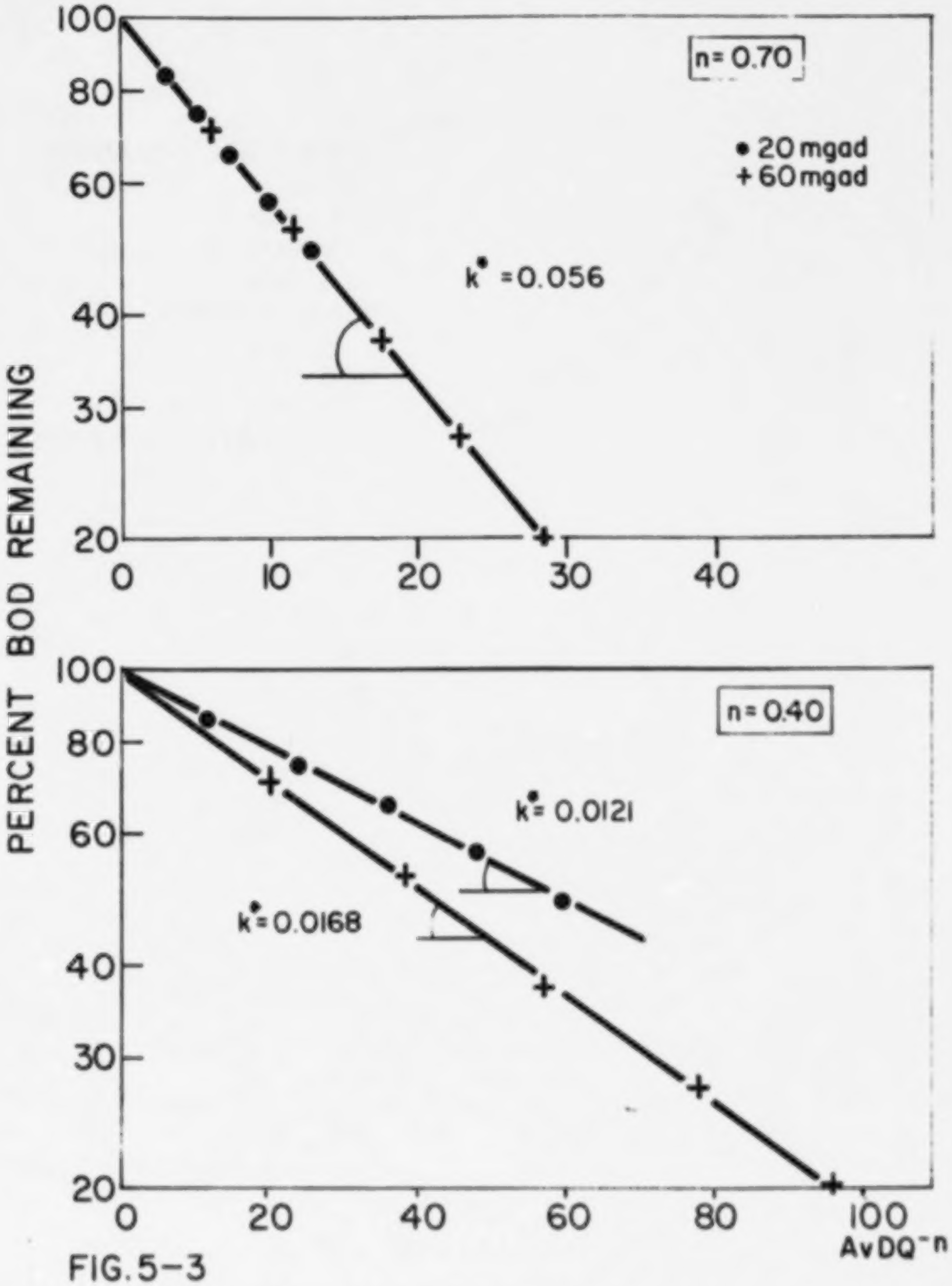

FIG. 5-3

INFLUENCE OF n ON k•

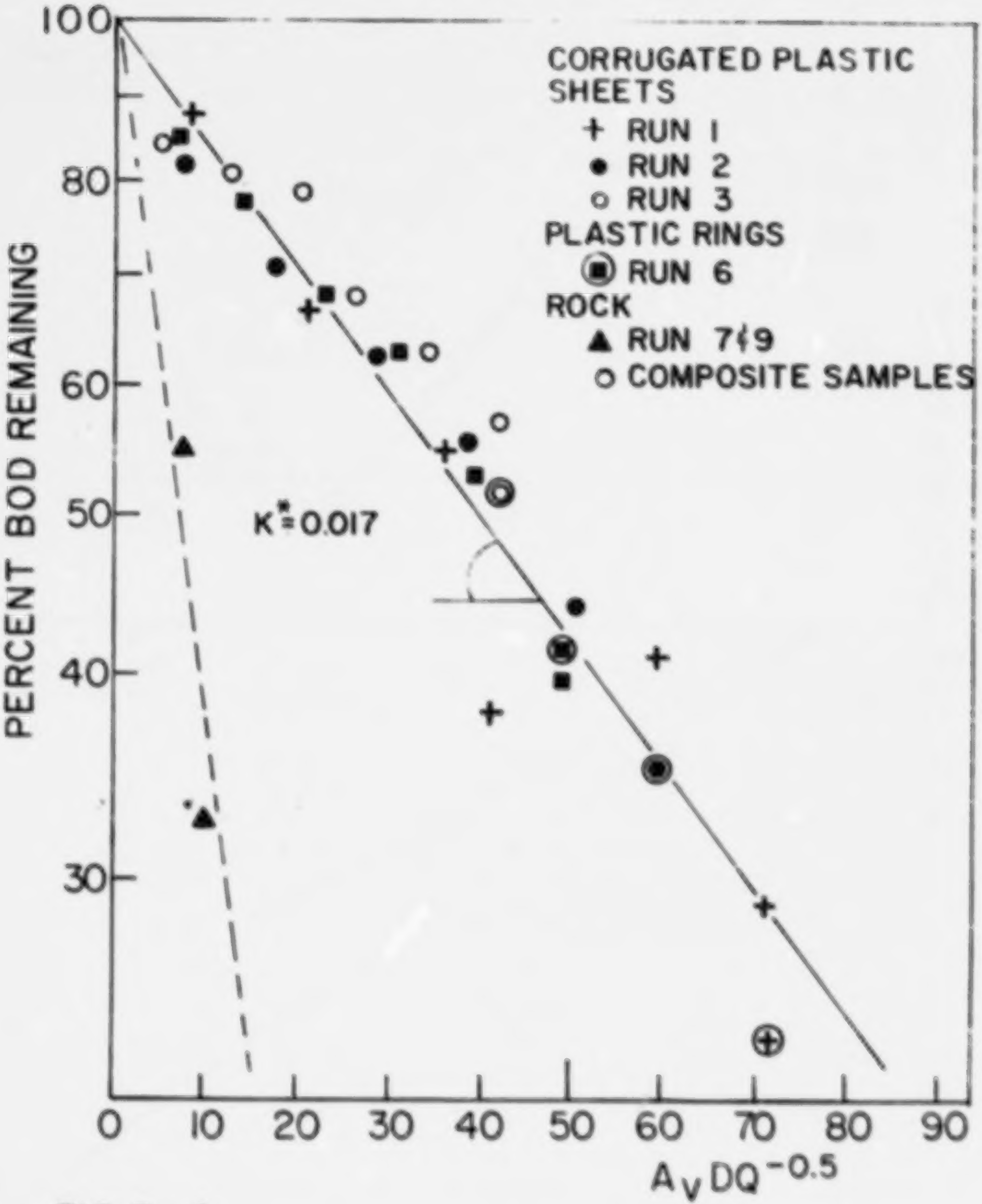

FIG. 5-4

DETERMINATION OF k*

PILOT PLANT DATA, GOVALLE, AUSTIN

fit. The average rate constant for the plastic media was k* = 0.017. The data observed for the rock filter medium result in a line which had a slope considerably different than those for the plastic media. The exponent n for corrugated plastic medium was evaluated using data from three experimental runs. The data reported by Gromiec and Malina (1970) indicate that n = 0.73. The value of the exponent n = 0.70 is used for evaluating the data observed when the filter medium was plastic rings. The data compare very well with the data observed for corrugated plastic sheets. The composite sample data observed for the rock medium will be comparable to the plastic medium data only if the exponent n = 0.36. These data are presented in Figure 5-5 and the slope of the line indicates that the rate constant k* = 0.050. This rate constant is approximately three times that calculated based on an exponent, n = 0.50.

These illustrations indicate the difficulty encountered when correlations of data from different experiments are attempted. The two procedures discussed have limited applications because the exponent, n, is affected by the hydraulic loading; however, the value does not reflect the effects of the detention time. The detention time in the trickling filter was expressed in Equation 5-3. The coefficient c_2 used in this equation can be assumed to be different for different filter media and is included in the overall rate constant, k*. Therefore, the rate constant can also be expected to vary with the type of filter medium.

The flow-through time is also a function of the configuration of the filter medium and the hydraulic surface load. Therefore, the flow-through time for different types of media can be characterized and the rate constant can be calculated independent of the detention time.

FACTORS AFFECTING PERFORMANCE

The overall performance of trickling filters traditionally has been related to the hydraulic surface loading and the organic loading. The performance could be correlated to either parameter when the concentration of BOD of the wastewater and the depth of the filter remain relatively constant.

The use of plastic media in place of rocks has led investigators to evaluate other parameters which relate to the performance of trickling filters. Attempts to correlate the quantity of filter slime to the actual surface area and in turn determine the organic load in terms of pounds of BOD per unit area of slime have been reported by Rincke and Wolters (1970) as well as by Lamb and Owen (1970). The effects of depth on the overall performance of trickling filters which were operated under identical conditions were reported by Keefer and Meisel (1952), Audion et al. (1970), Chipperfield (1967), and

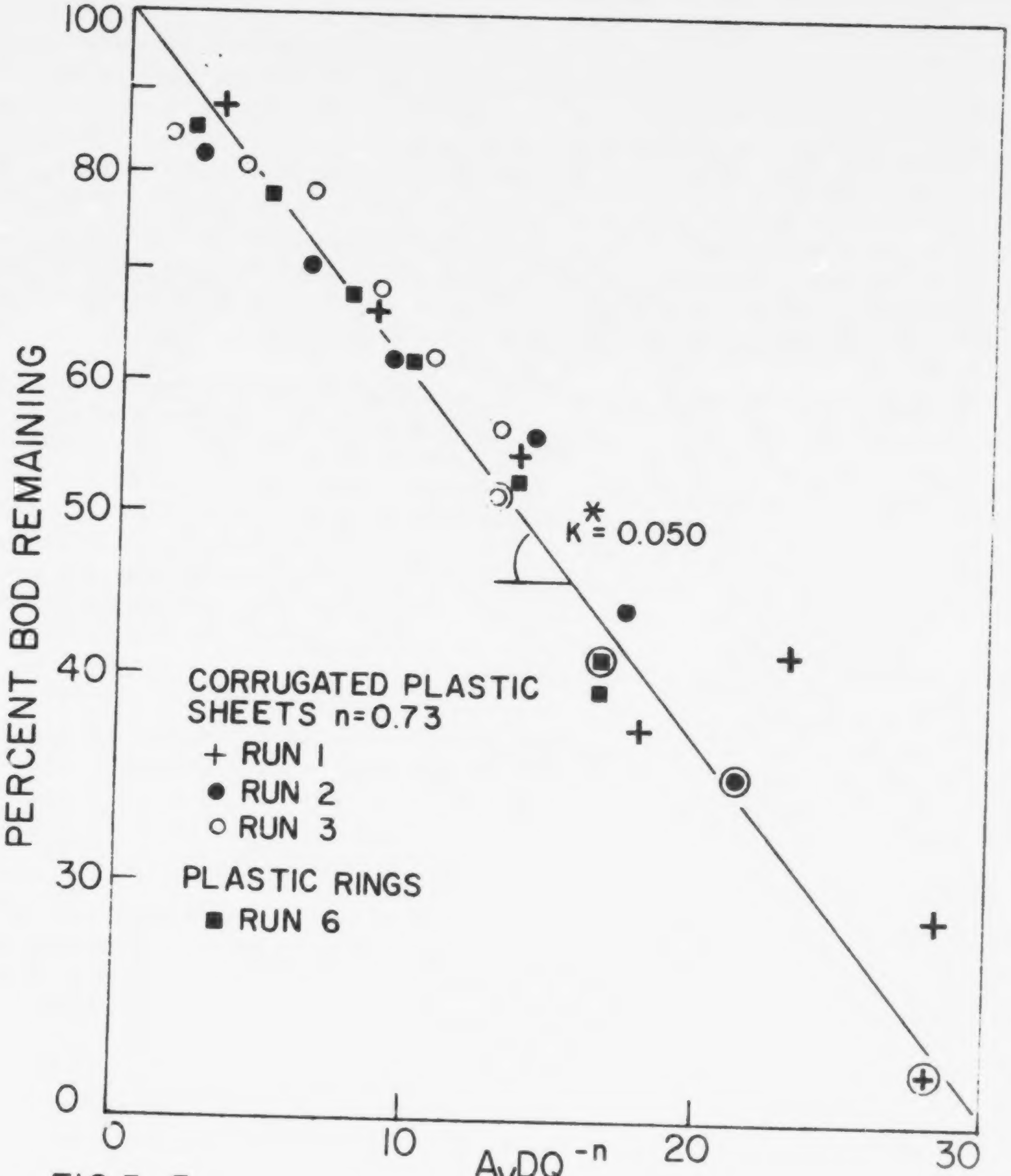

FIG.5-5
DETERMINATION OF κ^*
PILOT PLANT DATA: GOVALLE, AUSTIN

Bruce and Merkens (1970). Trickling filters were packed to a depth of up to 25 feet with corrugated or tubular plastic medium having specific surface areas of 12.2 to 67 square feet per cubic foot. The influent BOD was 200 to 400 mg/l. The results of these studies indicated that the volumetric BOD removal rate (lb BOD/1000 cu ft-day) was dependent only on the volumetric BOD loading (lb BOD applied/1000 cu ft-day) and was independent of the depth.

The relationship between BOD removal and the specific surface area is presented in Figure 5-6. The data indicate that the BOD removal increases as the specific surface area increases. The results of tracer studies conducted by Bruce (1970) are plotted in Figure 5-7. These data indicate that the relationship between the detention time in the filter and the specific surface area of the medium is approximately linear. The deviation from linearity of the data observed at higher specific surface areas may be partially attributable to the geometric configuration of the particular medium.

The performance of trickling filters is therefore related to the volumetric BOD loading as well as the specific surface area of the filter medium. The volume of filter medium required for effective treatment is considerably less for plastic media which have higher specific surface areas than for rock. Theoretically, the relationship between the volume required and the specific surface area may be presented as Equation 5-13.

$$\frac{V_x}{V_o} = \frac{A_{vx}}{A_{vo}} \qquad 5\text{-}13$$

in which

V_x = required volume of plastic filter medium (cu ft)

V_o = required volume of rock filter medium (cu ft)

A_{vx} = specific surface area of plastic medium (sq ft/cu ft)

A_{vo} = specific surface area of rock filter medium (sq ft/cu ft)

This relationship has not been completely verified; however, some of the results of recent experiments indicate that this relationship is valid. A safety factor to reflect the effective slime area is incorporated when filter medium other than rock is used. Equation 5-14 represents this relationship in which it is assumed that only 70 percent of the difference in the specific surface area of the plastic medium and the conventional rock medium is

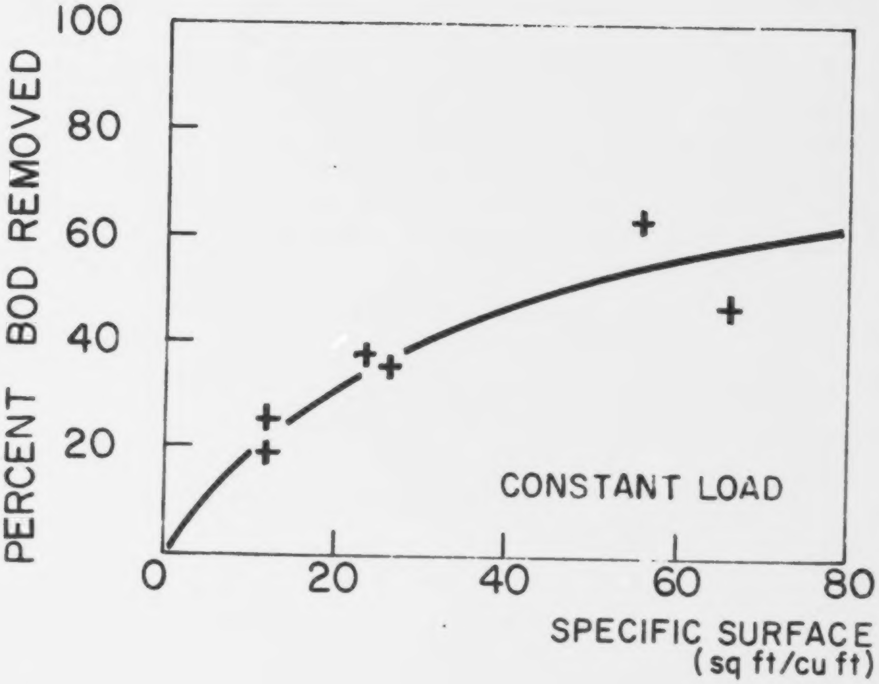

FIG.5-6
BOD REMOVAL VERSUS SPECIFIC SURFACE

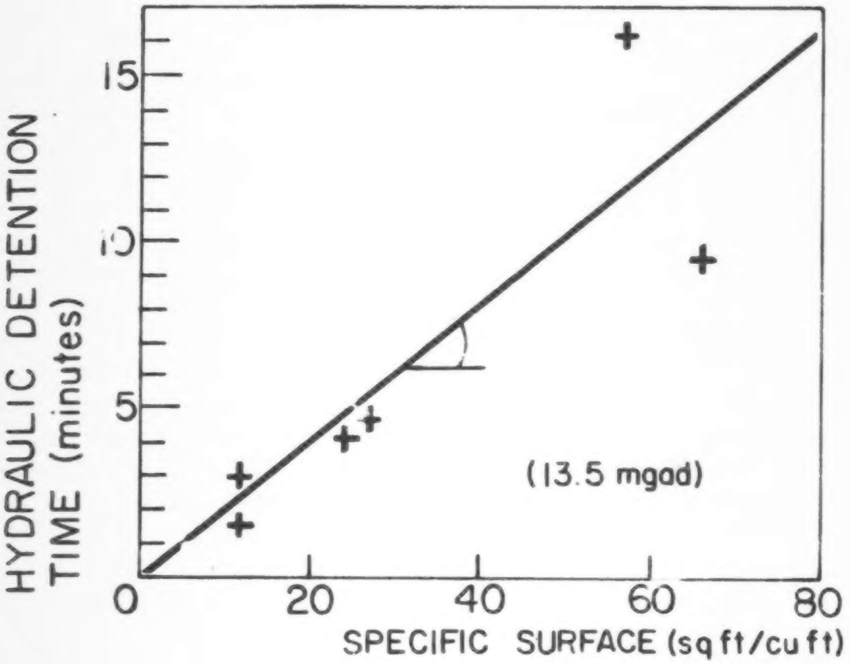

FIG. 5-7

HYDRAULIC DETENTION TIME VERSUS SPECIFIC SURFACE effective and that the specific surface area of rock filter medium, A_v = 12 sq ft/cu ft.

$$\frac{V_x}{V_o} = \frac{1}{1 + 0.7 \left(\frac{A_v - 12}{12}\right)} \qquad 5\text{-}14$$

The relationship between the fraction of volume required for plastic filter medium compared to conventional rock filter medium and the specific surface area are presented in Figure 5-8. The required volume can be reduced by 50 percent when filter medium which has a specific surface area of 30 sq ft/cu ft is used. Using a medium with a specific surface area of 60 sq ft/cu ft results in a filter which occupies only 25 percent of the volume required for conventional rock filter. However, increasing the specific surface area beyond 60 sq ft/cu ft does not markedly reduce the volume requirements. Therefore, the removal of BOD by trickling filters should increase as the specific surface area increases. The results of experiments by Bruce (1970) demonstrated this relationship.

The performance of trickling filters can be improved by increasing the specific surface area of the filter medium. This relationship is illustrated in Figure 5-9. The series of curves indicate that the efficiency of BOD removal of a high-rate rock trickling filter can be increased from 50 to 75 percent by replacing the rock medium with a plastic medium with a specific surface area of 60 sq ft/cu ft. However, as the efficiency of the rock filter increases, the improvement in the performance of the filter resulting from increasing the specific surface of the medium is less. For example, increasing the specific surface area of a high rate rock trickling filter operating at an efficiency of 95 percent would result in an efficiency of 98 percent. The specific surface area of 60 sq ft/cu ft seems to be the limiting value since negligible changes in the efficiency and performance of the filter are observed at higher values. Other data indicate that problems with clogging are more common when the filter medium has specific surface area of greater than 60 sq ft/cu ft.

Theoretically, the efficiency of BOD removal by low-rate trickling filters can also be increased by using a filter medium which has a higher specific surface area than rock. However, the results of investigations by Audion, et al. (1970) indicate that the increase in efficiency was negligible at various loading rates. This phenomenon may be explained by the fact that at low hydraulic loadings the slime surface is not uniformly wet; therefore, only a small portion of the slime is actually in contact with the wastewater. This uneven wetting is more noticeable when corrugated or tubular medium is used since the waste water flows vertically downward with very little lateral distribution. The results of tracer studies reported by Bruce and Merkens

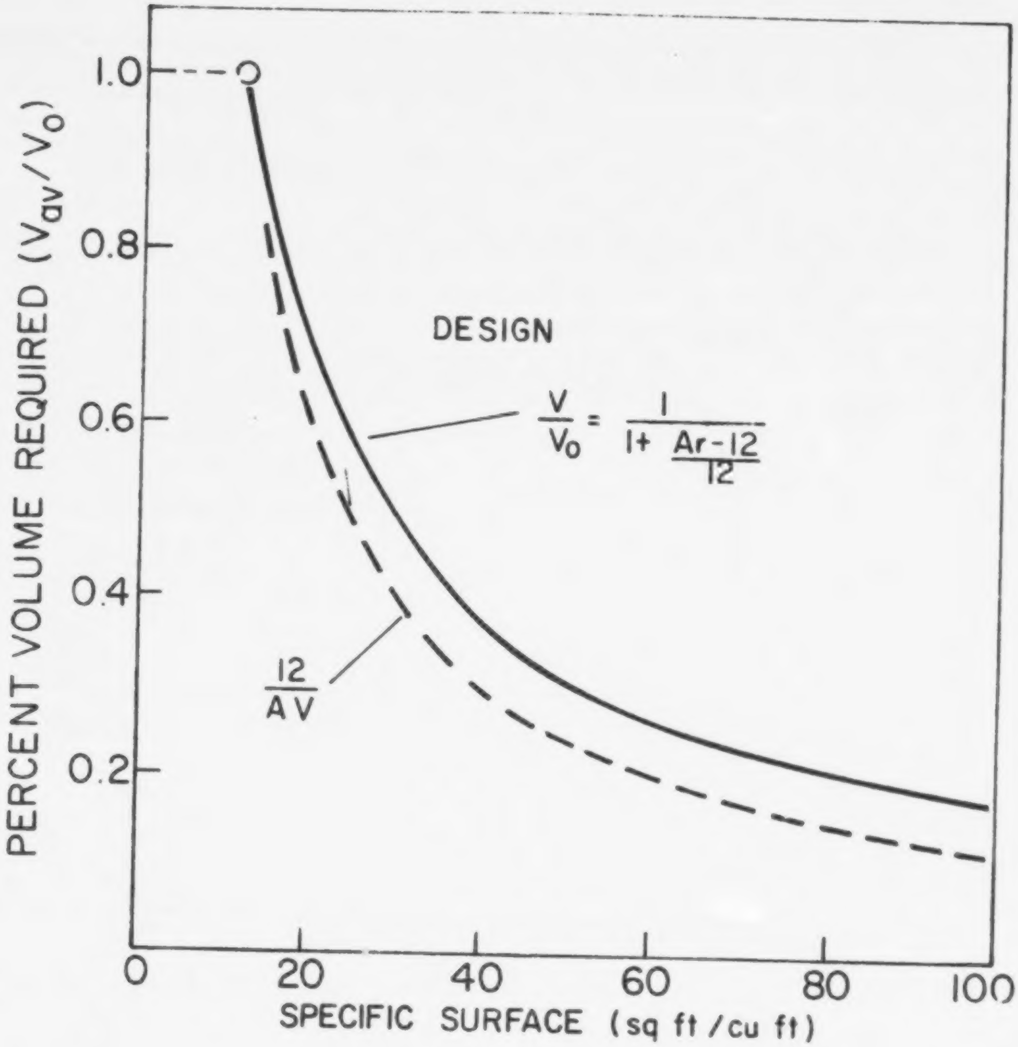

FIG. 5-8
RELATIONSHIP BETWEEN FRACTION OF VOLUME REQUIRED AND SPECIFIC SURFACE

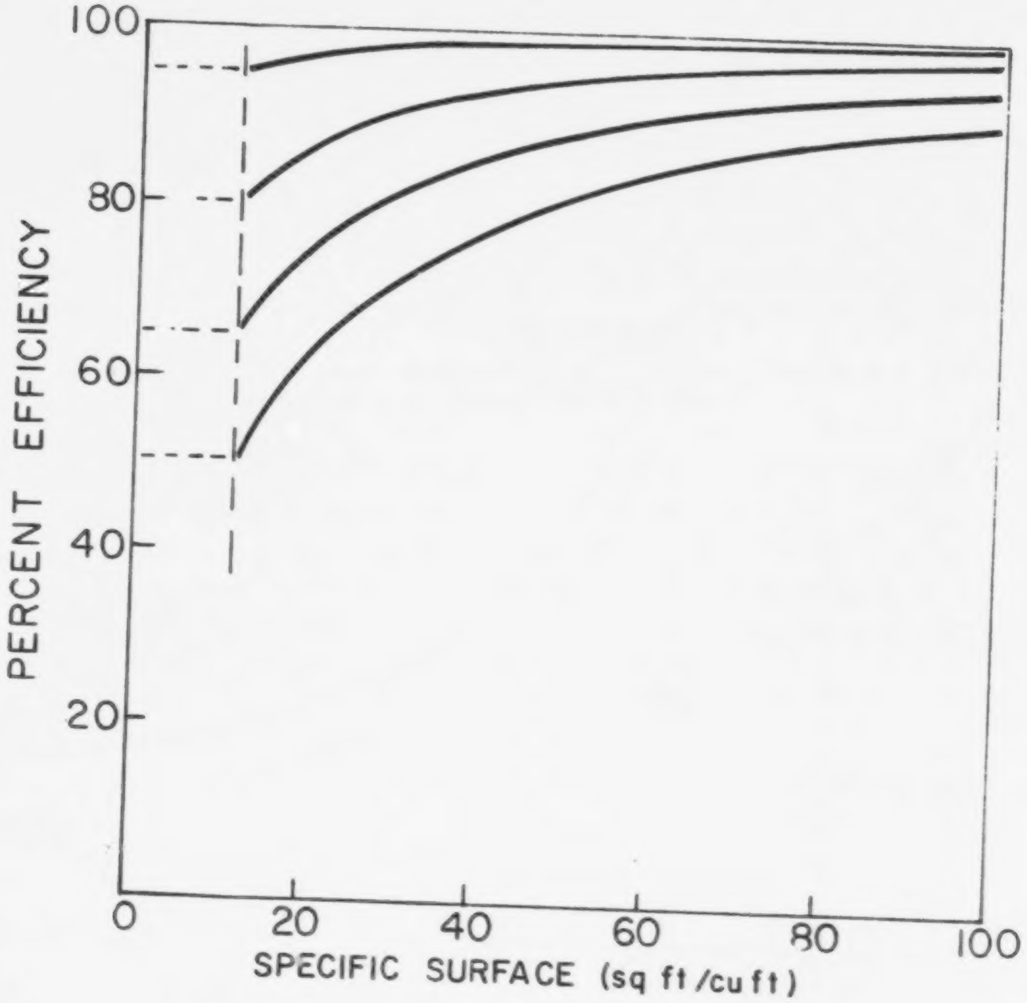

FIG. 5–9
PERFORMANCE AND SPECIFIC SURFACE

(1970) indicate that in a 7.5-foot deep trickling filter with corrugated plastic medium, about 90 percent of the tracer was recovered within an eight-inch diameter circle and that 95 percent of the tracer was recovered within a 16-inch diameter circle. These data indicate that uniform wetting of the surfaces of the filter medium is a very important design consideration. Bruce and Merkens (1970) indicated that the minimum hydraulic loading for a corrugated plastic medium for complete wetting of the surfaces was 10.5 gallons per square foot per hour.

The minimum hydraulic surface loading can be calculated based on the assumption that the wetting or the thickness of the liquid film varies linearly with the specific surface area. Equation 5-15 can be used to calculate the minimum required hydraulic service loading, based on the assumption that the specific surface area of rock medium $A_v = 12$ sq ft/cu ft and that the rock is uniformly wet at a hydraulic loading of 5 gal/sq ft-hour.

$$q_A = 5\frac{A_v}{12} = 0.42\,A_v \qquad 5\text{-}15$$

in which

q_A = hydraulic surface loading (gal/sq ft-hour)

A_v = specific surface area of filter medium (sq ft/cu ft)

The removal of soluble BOD is a function of the depth of the filter medium; however, the overall performance of a trickling filter was not affected by depth beyond some minimal depth required for the removal of soluble organics. The BOD of the effluent of the final clarifier following a trickling filter is a function of the concentration of suspended solids in the effluent. Bruce and Merkens (1970) reported that the nonsoluble BOD of the trickling filter effluent was equal to 60 to 80 percent of the concentration of suspended solids in the effluent. Therefore, the concentration of BOD in the effluent of a trickling filter when the suspended solids concentration is 30 mg/l would be at least 18 mg/l. Therefore, the overall performance of a trickling filter plant can be improved by increasing the efficiency of removal of suspended solids in the final clarifier. This relationship between the BOD and suspended solids in the effluent of biological processes has also been noted for the activated sludge process.

The minimum depth required for trickling filters should be based on the desired effluent concentration of soluble BOD. The significance of depth of the filter in connection with soluble BOD is illustrated in the following example. Equation 5-6 can be used to calculate the hydraulic surface loading for a 95 percent reduction in the concentration of soluble BOD for corrugated plastic filter medium at depths of ten and 20 feet, respectively. Other data include a rate constant, $k = 0.05$, specific surface area, $A_v = 27$ sq ft/cu ft and exponent, $n = 0.73$. Equation 5-6 can be reduced to the relationship between hydraulic surface loading and depth.

$$Q^{-0.73} = \frac{2.21}{D} \qquad \text{5-16}$$

The calculated hydraulic surface loading for a ten-foot and a 20-foot deep filter are 8.0 MGAD and 20.8 MGAD, respectively. Therefore, the volume of trickling filter required for 95 percent removal of soluble BOD for a 20-foot filter is approximately one-half the volume required for a ten-foot filter. Applying a hydraulic load of 20.8 MGAD to a ten-foot trickling filter would result in a soluble BOD removal efficiency of only 78 percent. Therefore, trickling filters should be designed to be deep enough to produce an effluent which has the desired concentration of soluble BOD. The overall performance of trickling filters is affected to some extent by recirculation of the effluent. Recirculation is recommended in high-rate filters in order to minimize clogging and ponding caused by growth of the slime. The problem of clogging is minimized to some extent by using corrugated plastic filter media which have a relatively high fraction of void spaces. The effect of recirculation on removal of soluble BOD can be determined by developing a mass balance, as indicated in Equation 5-17.

$$\frac{Q\,S_o + Q\,(RC)S_e}{Q\,(1 + RC)} = S_e \qquad \text{5-17}$$

The rate of recirculation is included in the term RC which is the ratio of the recirculated flow to the incoming flow. Equation 5-17 can be combined with Equation 5-6 and rewritten as

$$\frac{S_e^*\,[\,1 + (RC)]}{S_o^* + (RC)\,S_e^*} = e^{-kA_v D[(1 + RC)\ Q]^{-n}} \qquad \text{5-18}$$

$$\frac{S_e^*}{S_o^*} = \frac{e^{-kA_v D\,[\,(1 + RC)\ Q]^{-n}}}{[\,(1 + RC) - (RC)\,]\,e^{-kA_v D\,[(1 + RC)\ Q]^{-n}}} \qquad \text{5-19}$$

in which

S_o^* , S_e^* = concentration of soluble influent and effluent BOD (mg/l)

Q = hydraulic surface loading without recirculation (MGAD)

RC = Recirculation ratio = (rate of flow of recycle/rate of flow of influent)

The efficiency of BOD removal using recirculation can be calculated using Equation 5-19 and the parameters for corrugated plastic media. The efficiency for a ten-foot deep filter with a hydraulic surface loading without recirculation of 20.8 million gallons per day is approximately 78 percent. At a recirculation ratio RC = 1 and RC = 2 the efficiency was about the same, namely 74 percent. In this example, theoretically the performance of the trickling filter is not improved by using recirculation. Similar results were also reported for experiments using corrugated plastic medium at the Govalle Wastewater Treatment Plant in Austin, Texas. These data were observed at hydraulic surface loading rates of 94.5 MGAD without recirculation and with 100 percent recirculation. The effluent soluble BOD in both cases was approximately the same.

Recirculation can improve the soluble BOD removal efficiency of a trickling filter in those cases where the exponent, n, is relatively low and when the soluble effluent BOD is relatively high without recirculation. The quality of the effluent of trickling filters has usually been reported to be improved when recirculation is practiced. This improved effluent quality is more than likely the result of more uniform wetting of the filter medium and the result of washing out of the accumulated solids from the medium. Therefore, general design practice should include recirculation pumps for conventional (rock) trickling filters. The overall effect of recirculation on performance of a trickling filter plant should be evaluated for each particular installation.

The effluent of the trickling filter rather than that of the secondary clarifier should be recycled either directly to the filter or through the primary clarifier. This practice is preferred, since in many plants chemical flocculating agents would be added to the secondary clarifier in order to improve the efficiency of solids separation.

The operating temperature has an effect on the overall performance of a trickling filter plant in two ways. The biological processes are temperature dependent and the settling velocity of the suspended solids is affected by the change in viscosity resulting from temperature changes. The effect of temperature on the rate of biodegradation is expressed in Equation 5-20.

$$k_T = k_{20}\theta^{(T - 20)} \qquad 5\text{-}20$$

The temperature coefficient, θ, has been reported to vary from 1.035 to 1.072 (Eckenfelder, 1966). The discrepancy in the value of the temperature coefficient may be attributable to the difference in the depth of penetration of oxygen and the rate of oxygen uptake at different temperatures. Eckenfelder (1966) indicated that the effects of temperature changes would be much more marked at higher loading. Bruce (1970) indicated that the temperature coefficient, θ, for the overall performance of a trickling filter plant, ranged from 1.1 to 1.35 with an average value of 1.21. The average monthly temperature during these experiments varied from 9°C to 18°C (48°F to 64°F). During these studies, the efficiency in some of the filters dropped from 80 percent to 40 percent and in other filters from 55 percent to 20 percent as the temperature dropped. The effect of temperature on high-rate trickling filters must be considered in evaluating the performance of these treatment units.

ROCK TRICKLING FILTERS

The conventional trickling filter contains a rock medium which has a relatively narrow range of specific surface areas. Various mathematical models have been developed to describe the performance of the trickling filters in terms of hydraulic surface load or organic loading rather than specific surface area. Fairall (1966) developed the mathematical model expressed in Equation 5-21.

$$\frac{S_e}{S_o \; S_e} = \frac{1}{k} \; \frac{Q}{V} \qquad 5\text{-}21$$

This equation is illustrated graphically in Figure 5-10 in which the rate of substrate removal is related to the volumetric hydraulic loading. The slope of the line through the data is the reciprocal of the rate constant, 1/k. The wide range of the rate constant from k = 1 to k = 10 (day^{-1}) indicates that the rate constant is a function of the influent BOD or of the BOD volume load. The data presented by Bruce (1970) also indicate that the BOD volume load affects the rate constant. Therefore, the correlation of the BOD volume load to the efficiency of BOD removal is the preferred method of evaluating data.

The National Research Council (1946) compiled data for 34 operating trickling filters over an eight month period. The range of BOD removal was between 75 and 95 percent for organic loading of up to 2000 lb BOD/acre-day. These data are presented graphically in Figure 5-11 and the average performance of a trickling filter based on these data can be expressed by the model.

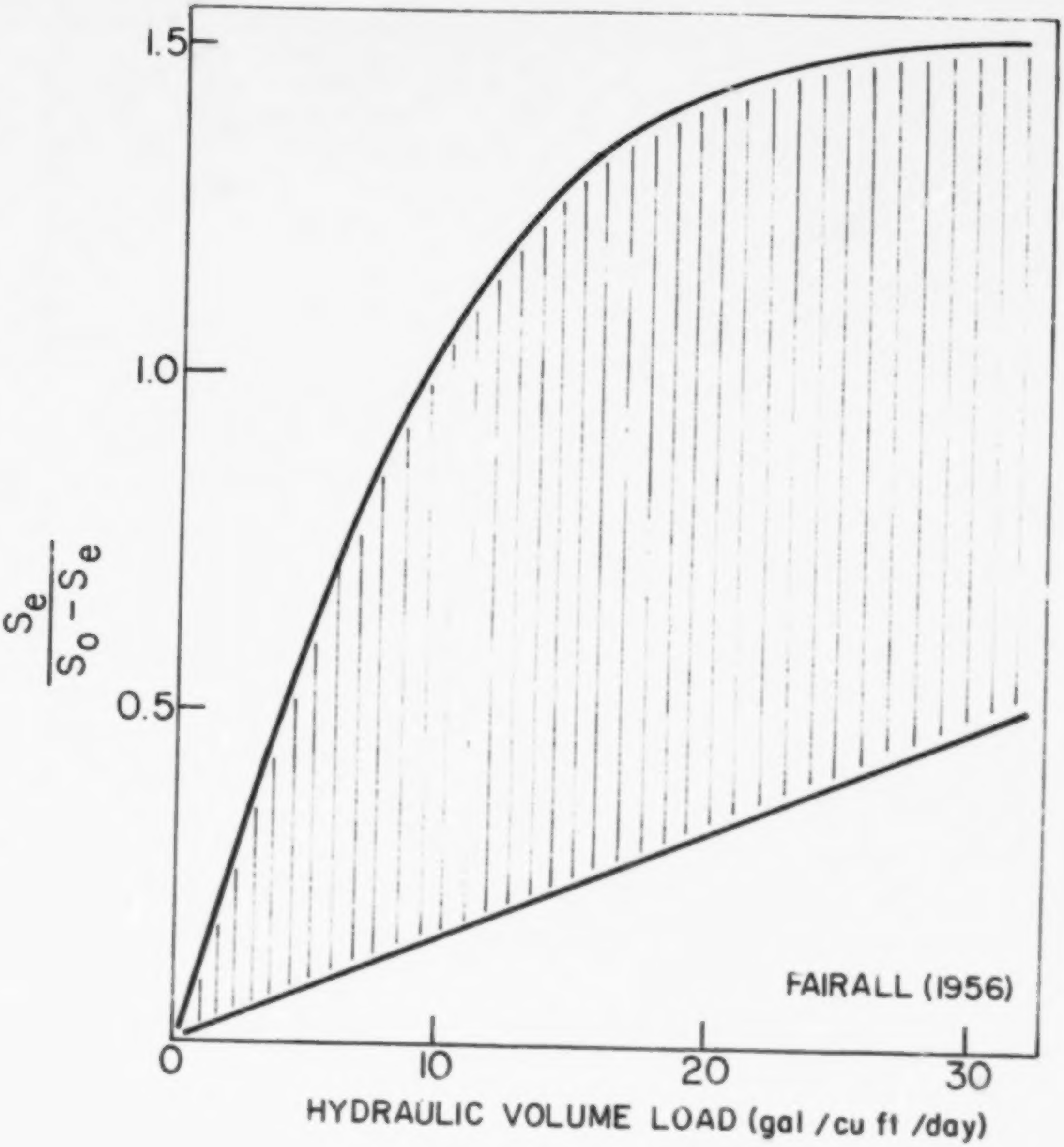

FIG. 5-10
RANGE OF RATE CONSTANTS

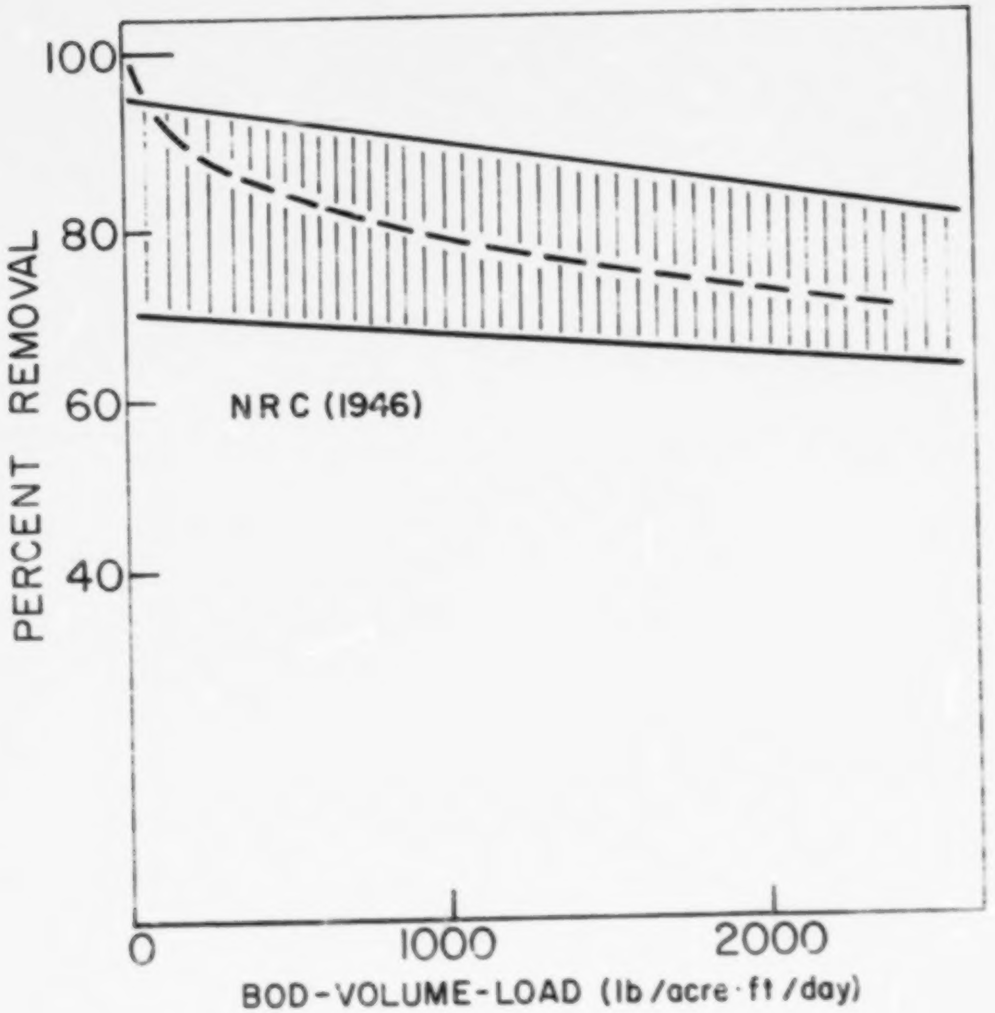

FIG.5-11
REMOVAL AND BOD VOLUME-LOAD

$$\left(\frac{S_o - S_e}{S_o}\right) = \frac{1}{1 + 0.0085\left(\frac{S_o Q}{VF}\right)^{0.5}} \qquad 5\text{-}22$$

in which

S_oQ = BOD applied (lb/day)

V = volume of medium (acre-feet)

F = number of effective passes $= \frac{1 + RC}{[1 + 0.1(RC)]^2}$

RC = recirculation ratio = (recycle flow rate/influent flow rate)

The effect of BOD volumetric load on the efficiency of trickling filters has also been reported by Keefer and Meisel (1952) and Rincke (1967). These data are presented graphically in Figure 5-12 and indicate that with increasing organic loading rates, the overall performance of trickling filters decreases. The average performance of trickling filters with depths of ten to 14 feet was developed by Rincke (1967) and is expressed in the form of the mathematical model:

$$100\left(\frac{S_o - S_e}{S_e}\right) = 93 - 0.272\,\frac{QS_o}{V} \qquad 5\text{-}23$$

in which

$\frac{QS_o}{V}$ = BOD volume load (lb BOD/1000 cu ft - day)

The NRC (1946) data cover a much broader spectrum of operating conditions than the data reported by Keefer and Meisel (1952) or by Rincke (1967). The data reported by Keefer and Rincke are much more comparable and cover a much narrower range, since these data were collected under relatively similar operating conditions.

EXCESS SLUDGE

The excess sludge produced in a trickling filter is a function of the rate of conversion of substrate to cell material, the endogenous respiration rate and the accumulation of non-biological solids in the filter slime. Sludge production

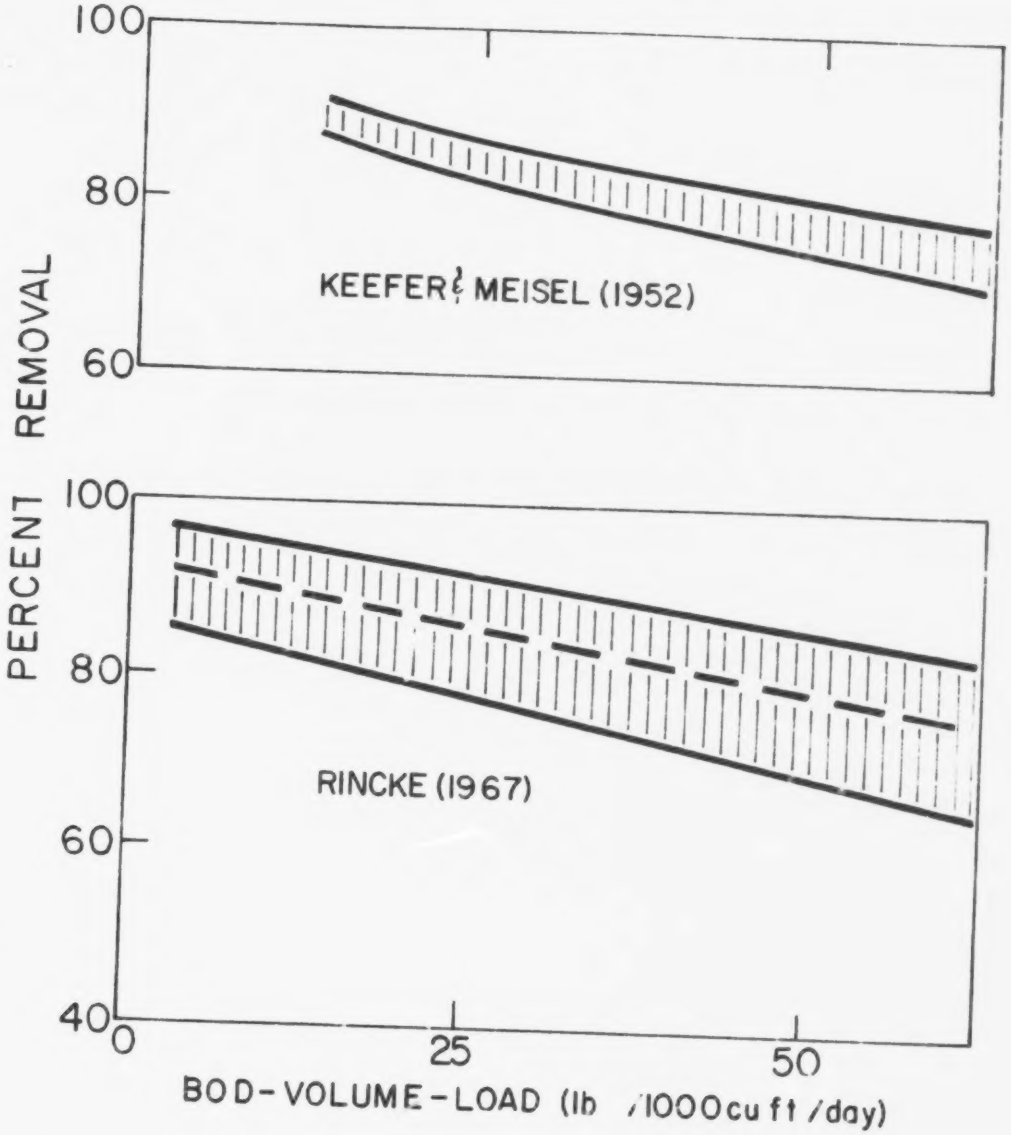

FIG.5-12
REMOVAL AND BOD-VOLUME-LOAD

in trickling filters is much similar to that developed in the activated sludge process. The quantity of excess sludge in a trickling filter plant can be determined by a mass balance presented as Equation 5-24.

$$QX_e = Q\,a\,(S_o - S_e^*) - bVX + QX_{on} \qquad 5\text{-}24$$

in which

Q = wastewater flow (MGD)

V = filter volume (cu ft)

X_e = effluent suspended solids from filter (mg/l)

X_{on} = influent nonbiodegradable suspended solids (mg/l)

S_o = influent BOD (mg/l)

S_e^* = effluent soluble BOD (mg/l)

X = biological sludge mass per unit volume (lb/cu ft)

a, b = constants

The sludge mass in a trickling filter can be related to the specific surface area by Equation 5-25.

$$VX = f\,(VA_v) \qquad 5\text{-}25$$

Combining 5-24 and 5-25 and rearranging the terms results in

$$\frac{Q}{VA_v}\,X_e = \frac{Q}{VA_v}\,[\,a\,(S_o - S_e^*) + X_{on}] - b \qquad 5\text{-}26$$

The constants in this equation can be evaluated graphically as illustrated in Figure 5-13 which represents a plot of the (BOD + SS) loading versus the excess sludge. The data are based on the assumptions that the effluent soluble BOD concentration is approximately zero and that the influent non-biodegradable suspended solids equal the total suspended solids entering the system. The data presented in Figure 5-13 represent experimental results reported by Bruce and Merkens (1970) and data observed at the Govalle Wastewater Treatment Plant in Austin, Texas, for rock and plastic ring media, respectively.

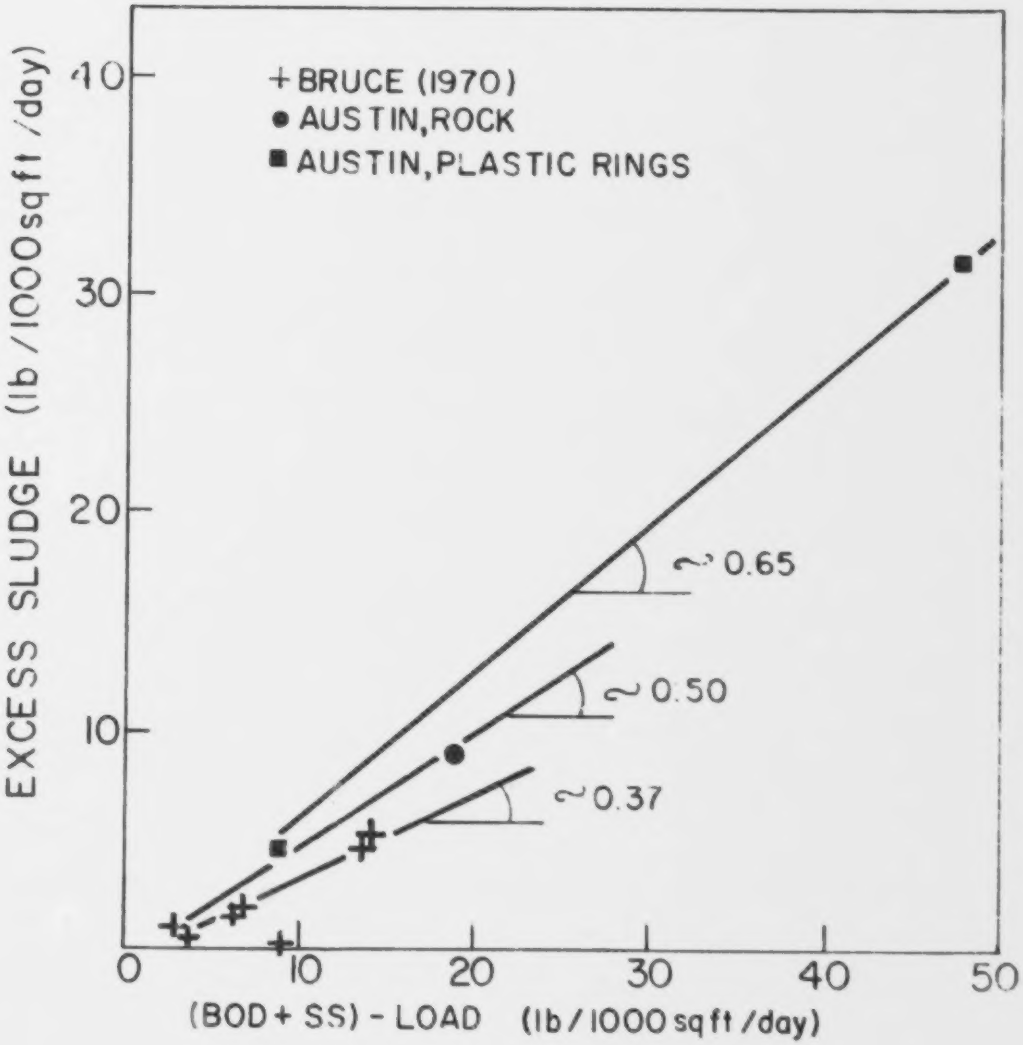

FIG. 5-13
EXCESS SLUDGE FROM TRICKLING FILTERS

The slope for the data reported for the Austin plant is somewhat higher than that for the other data. The wastewater at the Austin plant has a lower ratio of BOD to suspended solids which might account for the higher slope. The slope of the lines tend to decrease at the higher loading and is possibly the result of a lower rate of solids biodegradation, at the higher loadings than at lower loadings. An estimate of the excess sludge from a high-rate trickling filter can be generally determined by Equation 5-27 which is derived from the data plotted in Figure 5-13.

$$\frac{Q}{VA_v} X_e = 0.5 \frac{Q}{VA_v} (BOD + SS) - 0.5 \qquad 5\text{-}27$$

The excess solids are expressed in terms of pounds of solids per thousand square feet of filter medium per day in this equation. Equation 5-27 can be written as

$$X_e = 0.5 (BOD + SS) - 0.5 \left(\frac{VA_v}{Q}\right) \qquad 5\text{-}28$$

This equation indicates that 50 percent of the (BOD + SS) is converted to suspended solids which remain in the filter effluent. This conversion factor is somewhat lower than that reported for activated sludge systems.

The quantity of excess sludge produced in a low-rate trickling filter is much lower than that reported for high-rate filters or for the activated sludge process. The lower rate of solids accumulation may be attributable to the grazing activities of protozoa. The activity of the protozoa is reduced considerably at low temperatures. Therefore, slime tends to accumulate in the trickling filter during winter operation and the filter tends to unload the slime in the spring when the activity of the microorganisms is once again increased.

DESIGN FACTORS

The data required for design of trickling filter plants is the same information as is necessary for the design of activated sludge plants.

Conventional rock trickling filters require excellent solids removal in primary clarification to avoid clogging of the filter medium and ponding of the wastewater. Corrugated or tubular plastic materials provide a filter medium with relatively high void spaces; therefore, more influent suspended solids can be tolerated. A primary clarifier should precede any trickling filter treating municipal wastewater.

Recirculation of the effluent of the trickling filter to the primary clarifier or to the trickling filter may be practiced. The primary advantage of recirculation is a uniform wetting of the slime on the filter medium and possibly an increased removal of soluble BOD when high effluent concentrations result without recirculation.

Filter Volume and Filter Medium

The volume of the medium required and the cost can be calculated for trickling filters. This cost of rock can be compared with the cost of other filter media. The cost estimate should include the medium, the structure to house the medium, and the pumping requirements. A conventional rockline material has specific surface area which ranges between ten and 20 sq ft/cu ft whereas the plastic media have specific surface areas that range from 25 to 70 sq ft/cu ft. The filter medium can be classified as follows:

(a) randomly packed material (rocks, stones, plastic rings, raschid rings, etc.)
(b) self-supporting modules or corregated plastic sheets
(c) suspended corrugated sheets
(d) self-supporting plastic pipe or tubular materials

The main advantage of the plastic media is a high percentage of void spaces which is between 90 and 97 percent of the total volume compared to 30 to 50 percent of the volume for conventional rock filters. The void space provides for circulation of air as well as minimizes the potential clogging problems. As the specific surface area increases, the void space decreases.

The relationship between the clear space openings and the specific surface areas for various types and configurations of synthetic media are presented in Figure 5-14. It is interesting to note that square pipes, honeycomb pipes, spheres, and corrugated sheets which have the same diameter of open space or distance between sheets have about the same specific surface area. Circular pipes exhibit the highest specific surface area if the inside and outside surfaces of the pipe are included. However, at relatively small pipe diameters, the void space between the pipes becomes clogged with solids and the specific surface will decrease to about one-half the theoretical value. Corrugated plastic sheets have specific surface areas which are about twice that of a plain sheet. Calculation of the specific surface area for corrugated plastic sheets which are placed 2.1 inches apart indicate a theoretical specific surface area of 27.5 sq ft/cu ft. This value compares favorably with the value of 27 sq ft/cu ft reported by the manufacturers.

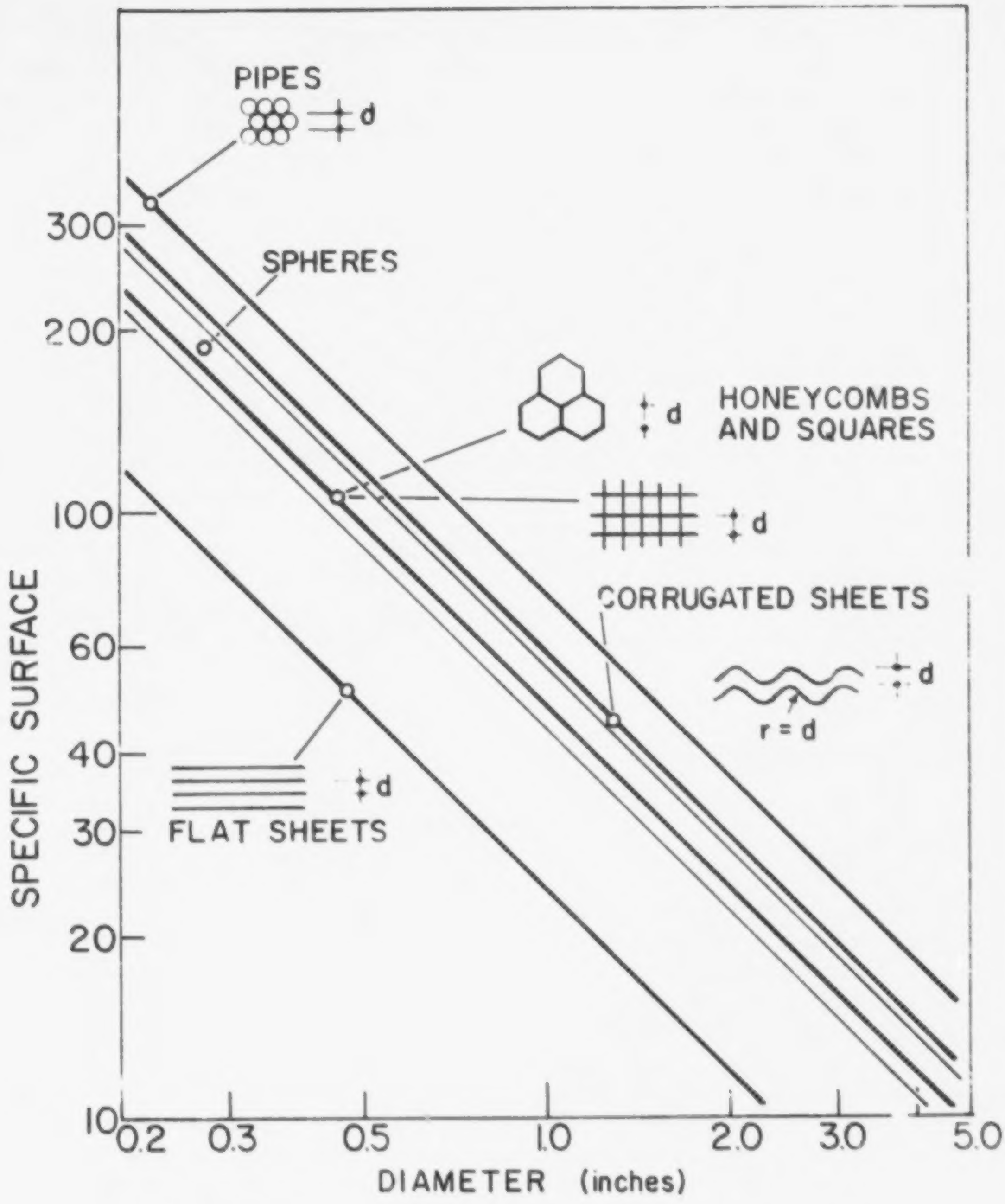

FIG. 5-14
RELATIONSHIP OF SPECIFIC SURFACE AND DIAMETER FOR VARIOUS MEDIA

Therefore, the curves in Figure 5-14 can be used to give preliminary estimates of the relationship of specific surface area and the distance between sheets or clear diameters. A clear opening of 0.7 - 0.8 inches is required to avoid clogging of the medium by slime. This distance therefore limits the maximum specific surface area to about 70 sq ft/cu ft.

Operating Factors

There are two operational techniques of improving the efficiency of a trickling filter system which is already installed. Recirculation may be employed or the dosing frequency can be changed. Recirculation can minimize ponding and clogging by washing out accumulated slime and solids. Hawks and Shepherd (1970) reported that the dosing frequency also can affect the operation of a trickling filter. Dosing frequencies of 0.3 and 14 minutes were used over a five-year period of operation. The results indicate that in the summer, the performance of the trickling filter was better at high dosing frequency and in the winter the performance was better at the low dosing frequency. At the lower dosing frequency, a high hydraulic loading is applied to relatively small areas of the filter and the washout effect is greater. At the low dosing frequency in the wintertime, less slime tends to accumulate than at the high dosing frequencies. Variation of dosing frequency might be considered to improve the operation of low-rate trickling filters as well.

Design Formulation

Various design formulations have been developed and various standards of design have been recommended. The Ten State Standards (1968) recommended design loading rates of up to 50 lb BOD/1000 cu ft - day. The BOD removal efficiency at this range of organic loadings is between 67 and 75 percent. These efficiencies are at the lower end of the range reported in Figures 5-9 and 5-10. These standards do not include any temperature effects, and it seems that the design performance includes winter conditions. Other published information indicates that the efficiency is limited to about 80 to 85 percent regardless of the type of filter medium and depth of medium used. Low-rate trickling filters can be operated at higher efficiencies than the high-rate trickling filters.

The results of studies conducted in England with smaller size graded rocks and slags indicate that detention times in these filters are relatively long and that high BOD removal efficiencies are possible. Eden (1964) reports that for a five-foot deep trickling filter, detention times of one to two hours are possible depending on the rate of hydraulic application. Therefore, a higher efficiency can be obtained at these longer detention times.

Trickling filters which are packed with plastic media require uniform wetting and the flow of the wastewater is essentially vertically downward. Therefore, long detention times are not possible with this type of medium. In England, two stage filtration is employed for many industrial wastes. The first-stage trickling filter is operated as a high-rate filter with a plastic medium. The second-stage trickling filter is operated at a low-rate using conventional rock or slag media.

The performance of single stage conventional rock filters at different BOD volume loads are presented in Table 5-2. The efficiency is expressed as the average as well as the recommended design values. At the lowest BOD volume loading, the effluent BOD concentration should average about 20 mg/l although values of 25 mg/l can be expected.

The required volume of filter medium can be reduced when plastic corrugated or tubular material is used. The reduction in volume required can be calculated using Equation 5-14, and the hydraulic surface load required to maintain uniform wetting of the medium can be computed from Equation 5-15. The excess sludge produced during high rate trickling filtration can be calculated from Equation 5-27.

TABLE 5-2

TRICKLING FILTER DESIGN

Conventional Filling Material
($A_v \sim 12$ sq ft/cu ft)

BOD Volume Load lb/10^3 cu ft/day	Efficiency Average	Efficiency Design
10	~82%	75%
50	~77%	67%
75	~70%	60%

WASTE STABILIZATION PONDS

PROCESS DESCRIPTION

Waste stabilization ponds are biological waste treatment systems in which algal photosynthetic and bacterial oxidation are effective in stabilizing a portion of the organic material in wastewaters. The process is controlled to a large extent by climatic conditions, primarily temperature and wind action.

A waste stabilization pond system can include a single pond, a number of ponds in series or parallel. The environment within the pond and the purpose for which the pond is used are the bases for classification of ponds as anaerobic ponds, facultative ponds and maturation ponds. These three types of ponds are generally used for the treatment of wastewaters. High-rate ponds have also been used in some cases for a growth of algae which are to be harvested.

Anaerobic ponds are those ponds in which no dissolved oxygen is detected immediately below the surface. The rate of oxygen utilization is much greater than the rate of reoxygenation. Anaerobic ponds for the treatment of municipal wastewater are essentially large sedimentation and digestion basins. Long detention times are provided for the settled solids to undergo anaerobic degradation. The effluent of an anaerobic pond contains a relatively high concentration of soluble and colloidal organic material (BOD) and requires additional treatment.

Facultative ponds are most commonly used for the treatment of wastewaters. A benthic anaerobic zone of activity is overlain by an aerobic zone of biological activity near the surface. A portion of the suspended solids entering a facultative pond will settle to the bottom and undergo anaerobic decomposition. The soluble organic material in the wastewater and those soluble organics released by the anaerobic degradation of the organic solids are used by heterotrophic bacteria and converted to carbon dioxide and new bacterial cellular material. The oxygen required for aerobic activity is provided by algal photosynthesis and to a lesser extent by diffusion of atmospheric oxygen into the surface waters of the pond. The effluent of facultative ponds will contain a relatively low concentration of soluble BOD and varying concentrations of algal cells.

Maturation ponds are used to polish the effluent of facultative ponds or of other biological treatment systems. A reduction in the bacterial content,

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suspended solids, and some nutrients are realized in maturation ponds. The removal of BOD in maturation ponds is negligible.

KINETIC MODELS AND DESIGN CONSIDERATIONS

The number of mathematical models developed for the design of waste stabilization ponds is limited, although pond systems have been widely used and the operation of ponds investigated during the past twenty years. Most of the mathematical models have not been developed in sufficient detail to completely describe the reactions and microbial activity within the pond. In other cases, the application of the mathematical models is limited to a specific geographical region or to ponds of a specific design.

ANAEROBIC PONDS

The mechanism of anaerobic degradation of organic material involves the sequential activity of facultative microorganisms and methane forming bacteria. Complex organic material is hydrolyzed by the facultative bacteria to organic acids, which can be converted by methane forming bacteria to methane and carbon dioxide gases. Therefore, carbon can be removed from the system by methane fermentation.

Anaerobic ponds are designed to maintain environmental conditions which are favorable for the development of methane bacteria. The primary factors affecting the growth of methane bacteria are temperature, pH, detention time, and organic loading rate. Methane bacteria grow relatively slowly compared to facultative organisms and require much longer detention times for development of an adequate population to carry out effective methane fermentation. The presence of free dissolved oxygen in the environment can be inhibitory to the methane bacteria. The facultative bacteria use the dissolved oxygen and protect the methane bacteria from exposure to oxygen. However, a thin algal surface layer in a pond will help minimize odors. The minimal temperature for active growth of methane bacteria is approximately 20°C. However, methane bacteria will continue to grow but at a much reduced rate at temperatures as low as 15°C, but the rate of gas production will become almost negligible. The pH range for effective methane fermentation is between pH 6.6 and pH 7.2.

The hydraulic detention time in anaerobic ponds should be between two and five days and is about the same as the generation time of the more rapidly growing methane forming bacteria. Other methane bacteria have generation times of 20 to 30 days. The more slowly growing methane forming bacteria are more likely to be found in the sludge deposits in the anaerobic pond than in the liquid phase. At these long hydraulic detention times most of the suspended solids will settle to the bottom. The organic fraction of the solids in the benthic zone will undergo decomposition and some of the organic material will be converted to methane gas. Aguirre and Gloyna (1971) indicate that a 50 percent removal of BOD was observed for anaerobic ponds treating municipal wastewater at the Govalle Treatment Plant in Austin, Texas, at organic loadings of 900 pounds of ultimate BOD per acre per day (six lb BOD_u/1000 cu ft - day) and a detention time of 2.3 days at a temperature of 16°C (60°F). No odors were noticeable at these conditions. However, at an organic loading of 1600 lb BOD_u/acre-day (11 lb BOD_u/1000 cu ft-day) and a detention time of 1.3 days, considerable odors were noticeable although the removal of BOD was also approximately 50 percent. At the higher loading, the temperature was 28°C (82°F), and the production of odor was probably the result of active fermentation and mixing rather than inadequate detention time.

White (1970) indicated that the average volumetric loading recommended in various states was 12 to 15 lb BOD/1000 cu ft - day. However, the volumetric loadings varied from a minimum of three lb BOD/1000 cu ft - day for Georgia to a maximum of 1000 lb BOD/1000 cu ft - day in Texas. The Texas regulations also recommend that the solids loading to the pond be between 100 and 400 lb SS/1000 cu ft - day and detention time between five and 30 days. A comparison of these recommended criteria with the results reported by Aguirre and Gloyna (1971) indicate that the higher loadings are for the treatment of industrial wastes in which the concentration of organic material and settleable solids are relatively high, and high loading rates are possible at relatively long detention times. However, with domestic wastewater in which the concentration of BOD is approximately 200 mg/l loadings of 12 to 15 lb BOD/1000 cu ft - day would be achieved at hydraulic detention times of one to two days.

The depth of anaerobic ponds should be between 10 and 15 feet. Sufficient volume should be provided for the storage of sludge resulting from the accumulation of settled solids. The introduction of the incoming wastewater near the bottom of the pond and distributed as uniformly as possible over the entire area of the pond is preferred.

Facultative Ponds

Facultative ponds are waste stabilization systems in which the removal of organic material takes place via two mechanisms, namely anaerobic degradation and facultative or aerobic biodegradation of the organic material by bacteria with oxygen provided by photosynthesis by algae. The accumulated organic solids in the sludge layer are degraded anaerobically and result in the formation of methane, carbon dioxide and soluble compounds.

Algal photosynthesis produces some of the oxygen required to maintain aerobic, odor-free conditions. Algae utilize the inorganic carbon which is made available from the aerobic respiration, the anaerobic decomposition of organic solids, the atmosphere and the influent wastewater. This carbon is converted to cell material, and as such, is removed only when the algae are harvested.

Aerobic respiration is responsible for the conversion of the organic material to new cell material as well as the carbon dioxide which is then available for algal photosynthesis. For facultative ponds the concentration of soluble BOD of domestic wastewater under favorable environmental conditions may be reduced to a residual level of approximately 15 mg/l.

The composition of algal cells includes carbon, nitrogen and phosphorus in a ratio of about 106:16:1 expressed as millimoles per liter (Stumm and Morgan, 1963). The aerobic utilization of one mole of organic carbon results in the conversion of approximately 65 percent of the carbon to bacterial cell mass and approximately 35 percent of the carbon is released as carbon dioxide (Wuhrmann, 1964). In addition to the utilization of the organic material, approximately 0.65 moles of carbon dioxide are produced per mole of oxygen uptake. Algae release one mole of oxygen for each mole of carbon dioxide utilized. Therefore, a net deficit of 0.35 moles of carbon dioxide exist. This deficit is in part satisfied by the anaerobic degradation of cell material. If the dead cells undergo aerobic degradation, all of the carbon is released as carbon dioxide and re-enters the carbon cycle in the pond. However, since the dead cells generally settle to the bottom of the pond and undergo anaerobic decomposition, about 70 percent of the carbon is released as methane gas and 30 percent of the carbon re-enters the carbon cycle, as carbon dioxide.

Two of the most important factors in facultative waste stabilization pond design include:

(a) the dependence of algal growth on the total carbon in the system
(b) The primary source of carbon is also derived from methane fermentation

Therefore, the theoretical design of waste stabilization bonds should be oriented toward a balanced system in which the algal population is maintained at the minimal level to insure aerobic conditions in the liquid zone. However, it is practically impossible to control algal production at either a maximum or minimum because of changes in the environment in the pond as well as in the available nutrients. Therefore, the design should be based on maintaining a sufficient algal population to minimize anaerobic conditions and to provide for efficient operation. The algae in the effluent of a properly designed pond will probably represent 90 percent of the total BOD leaving the pond. However, these cells will not exert an immediate oxygen demand on the receiving stream, but do represent a suspended solids load.

The various processes involved in the efficient operation of a facultative stabilization pond are illustrated schematically in Figure 6-1. The interaction among the anaerobic decomposition in the benthic zone, the aerobic respiration and the algal photosynthesis are shown.

The primary design criteria for facultative ponds involve maintaining a minimum soluble organic concentration as well as a minimum algal concentration in the pond effluent. The bacterial population developed in a pond can be calculated from a mass balance illustrated as Equation 6-1 and modified as Equation 6-2.

$$QX_a = a\,Q\,(S_o^* - S_e^*) - bVX_a \qquad 6\text{-}1$$

$$X_a = \frac{a\,(S_o^* - S_e^*)}{1 + bt} \qquad 6\text{-}2$$

in which

X_a = concentration of heterotrophic bacteria (mg/l)

S_o^* = concentration of soluble BOD in the influent plus the BOD released by anaerobic decomposition (mg/l)

S_e^* = concentration of soluble BOD in effluent (mg/l)

a = coefficient of conversion of BOD removed to bacterial cells

b = coefficient endogenous respiration rate (day^{-1})

t = coefficient of detention time of liquid in aerobic zone of a pond (day)

These equations are based on the assumption that the aerobic portion of the ponds is completely mixed and that the removal kinetics of soluble organic material are similar to those of the aerated lagoon process. The model

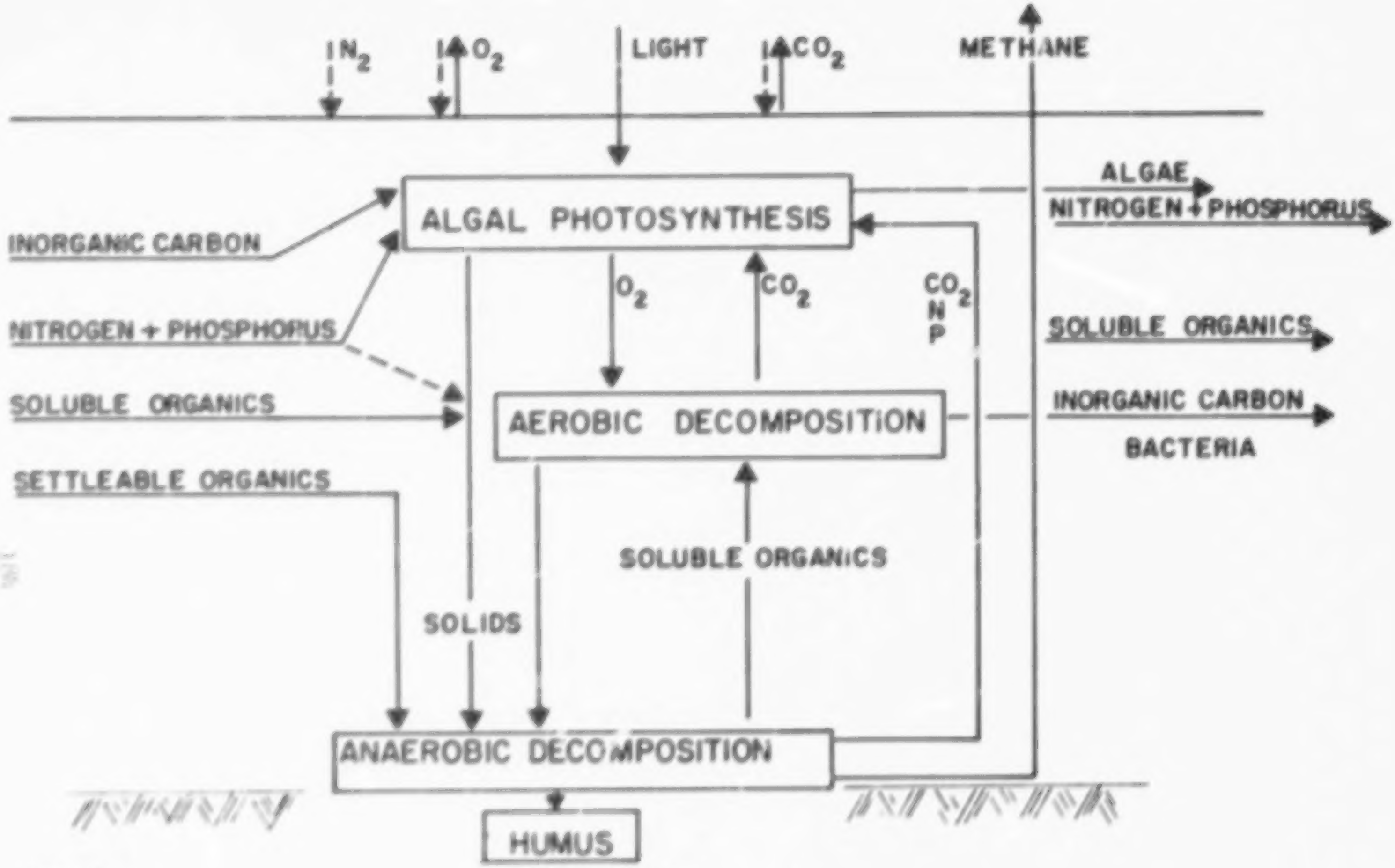
IN$_2$
O$_2$
LIGHT
CO$_2$
METHANE
ALGAL PHOTOSYNTHESIS
INORGANIC CARBON
NITROGEN + PHOSPHORUS
SOLUBLE ORGANICS
SETTLEABLE ORGANICS
O$_2$
CO$_2$
CO$_2$ N P
AEROBIC DECOMPOSITION
SOLUBLE ORGANICS
SOLIDS
ANAEROBIC DECOMPOSITION
HUMUS
ALGAE
NITROGEN + PHOSPHORUS
SOLUBLE ORGANICS
INORGANIC CARBON
BACTERIA

FIG.6-1
MECHANISMS OF DEGRADATIONS IN FACULATIVE PONDS

for the biological degradation of soluble organics in the completely mixed reactor is presented in Equation 6-3.

$$\frac{S_o^* - S_e^*}{X_a t} = kS_e^* \qquad 6\text{-}3$$

in which

k = BOD removal rate constant

The soluble organic material in the effluent can be calculated by combining Equations 6-2 and 6-3.

$$S_e^* = \frac{1 + b\,t}{a\,k\,t} = \frac{1}{a\,k\,t} + \frac{b}{a\,k} \qquad 6\text{-}4$$

This mathematical model indicates that a residual soluble organic concentration ($S_e^* = b/ak$) will remain in all systems. The concentration of organic material in the effluent above this residual is a function of detention time as well as the reaction rate constant. In this particular model, the constants b and k are temperature dependent, and can be calculated for any temperature by using either Equation 6-5 or 6-6.

$$k_T = k_{20}\theta^{(T-20)} \qquad 6\text{-}5$$

$$b_T = b_{20}\theta^{(T-20)} \qquad 6\text{-}6$$

The mathematical model implied bacterial growth. The concentration of heterotrophic bacteria in the pond will be higher as the rate of substrate removal $[(S_o^* - S_e^*)/t]$ increases. This model can be applied to the entire pond or to sections of the pond. At the inlet section the influent substrate (S_o^*) is relatively high, and the bacterial activity will be higher than at the outlet section where the substrate concentration is relatively low. However, the gross removal of soluble substrate will be dependent upon the volume of the aerobic zone and the duration of aerobic conditions. For example, if the depth of the aerobic zone of a pond during the daytime hours is one half the pond depth and if the dissolved oxygen concentration in the pond is zero during the night-time hours, the volume of the aerobic zone would be one fourth the volume of the entire pond and the duration of aerobic conditions would be only one fourth of the hydraulic detention time. The aerobic zone can be represented as a specific volume of the pond (V'). Therefore, the soluble effluent BOD should be a function of the organic load applied to the aerobic volume and the temperature of the pond. This relationship can be written as Equation 6-7.

$$S_e^* = f\left(\frac{QS_o}{V'}\right), (T) \qquad 6\text{-}7$$

However, the depth of the aerobic zone will fluctuate daily and seasonally. Therefore, a more practical relationship includes the effluent BOD and BOD surface load (lb BOD/acre) as shown in Equation 6-8.

$$S_e^* = f\left(\frac{QS_o}{A}\right), (T) \qquad 6\text{-}8$$

Hermann and Gloyna (1958) proposed one of the first rational design equations for stabilization ponds. A modified presentation of this mathematical formulation is presented as Equation 6-9, which can be used to calculate the required pond volume for the removal of 90 percent of the influent BOD and is related to the fixed optimum detention time (t_o) at 35°C. The optimum detention time is based on an average influent BOD of municipal wastewater, $S_o = 200$ mg/l. This relationship also includes a factor to compensate for any algal toxicity which might result from constituents in the municipal wastewater (Hwang and Gloyna, 1967) and a sulfide compensation factor (Espino and Gloyna, 1968).

$$V = K\left(\frac{QS_o}{200}\right)(t_o)\,\theta^{(35 - T)}\,(f)\,(f') \qquad 6\text{-}9$$

in which

V = pond volume required (acre - ft) or (cu ft)

Q = wastewater flow rate (gal/day)

S_o = influent BOD concentration (mg/l)

t_o = optimum time (days)

θ = temperature coefficient

T = minimum monthly temperature (°C)

f = algal toxicity or compensation factor

f' = sulfate, SO_2 effect ($f' = 1$ for $SO_2 < 500$ mg/l as SO_4)

K = conversion factor to make units of the terms compatible

Marais and Shaw (1961) applied first order kinetics to the waste stabilization pond performance and assumed completely mixed conditions. The mathematical model which was developed is presented as Equation 6-10.

$$\frac{S_e}{S_o} = \frac{1}{1 + kt\theta^{(35-T)}} \qquad 6\text{-}10$$

in which

S_o, S_e = influent and effluent BOD (mg/l)

k = rate constant (day^{-1})

t = hydraulic detention time (day)

These mathematical models provide a reliable foundation for estimating the required size of waste stabilization ponds. In general, these equations consider the influence of temperature and the biodegradation rate. Marais and Shaw (1961) recommend a value of the rate constant, $k = 0.17$ which was calculated from data observed in full-scale stabilization ponds. Hermann and Gloyna (1958) applied an optimum detention time (t_o) developed from laboratory-scale work in baffled aquaria.

The relationship between the percent BOD remaining and detention time for the Hermann-Gloyna equation are illustrated in Figure 6-2. The data reported by Suwannakarn and Gloyna (1963) were for ponds operated at a temperature of 35°C and the substrate was synthetic wastewater; however, the data reported by Hermann and Gloyna were for municipal wastewater. The optimum detention time for 90 percent BOD removal was about 3.8 days. The optimum rate constant, $k_{35} = 0.60$, was determined graphically from Figure 6-2. This rate constant can be adjusted to a temperature of 20°C using the relationship shown as Equation 6-5 and the rate constant $k_{20} = 0.18$ is for a temperature coefficient, $\theta = 1.085$. This rate constant is very similar to that reported by Marais and Shaw (1961) for a temperature of 20°C. The temperature coefficient, θ, for stabilization ponds was developed by Suwannakarn and Gloyna (1963). The results of these investigations are presented in Figure 6-3 in which the effect of temperature on BOD removal is illustrated.

The optimum detention time for a single pond can be calculated from Equation 6-10 based on the assumption that the rate constant is the same in plug flow and completely mixed systems. The optimum detention time (t_o) therefore is 15 days at an operating temperature of 35°C based on the assumption that

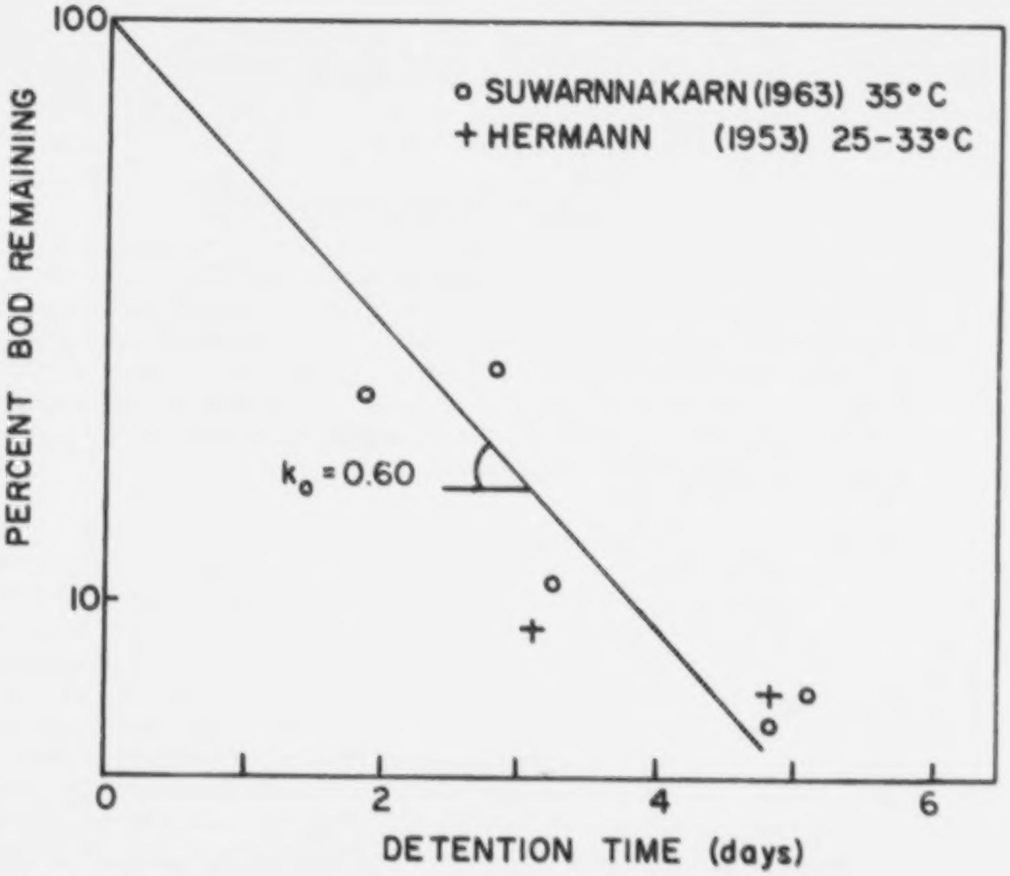

FIG. 6-2
BOD REMOVAL IN LABORATORY PONDS

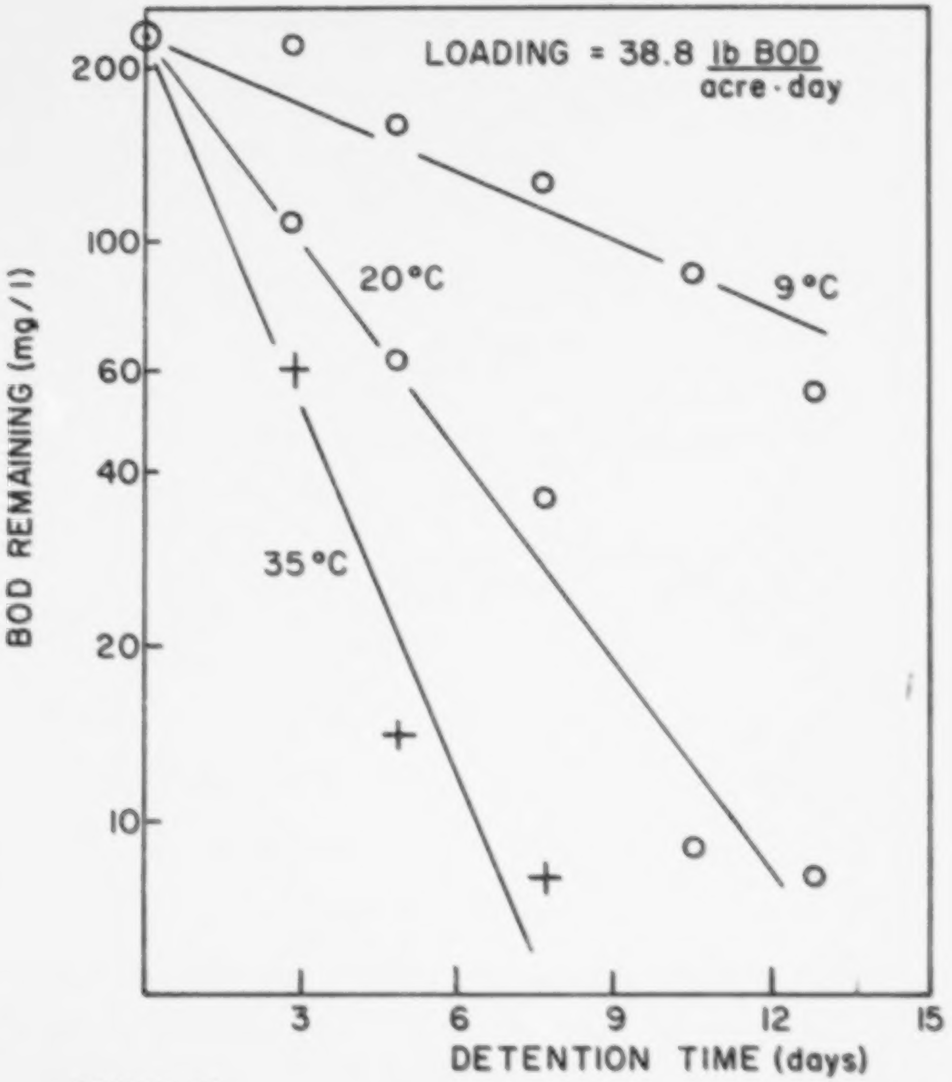

FIG.6-3

BOD REMAINING IN LAB SCALE PONDS (SUWARNNAKARN – 1963)

90 percent of the soluble BOD is removed; therefore, $S_e/S_o = 0.1$ and on a rate constant $k = 0.60$. This apparently long detention time may be explained by the fact that the equation is representative of a completely mixed system while the rate constant was based on data observed in a plug flow system. The substrate used in the experiments reported by Suwannakarn (1963) was almost completely soluble and readily biodegradable.

The hydraulic regime established in most stabilization pond systems is somewhere between a plug flow and completely mixed systems. Therefore, the recommended optimum detention time (t_o) of seven days was introduced by Gloyna (1969) as a modification of the original relationship. This optimum time is applicable to those systems which closely approximate the conditions in rectangular stabilization ponds with length to width ratios of between 2:1 to 3:1.

Equation 6-9 can be modified and rearranged in terms of areal loading

$$V = \frac{Q}{0.326} \cdot \frac{S_o}{200} (t_o)\, 1.085^{(35-T)} \qquad \text{6-12}$$

$$\frac{QS_o\,(8.34)}{A} = \frac{0.326\,(200)\,(8.34)\,d}{1.085^{(35-T)}\,(t_o)} \qquad \text{6-13}$$

$$\frac{Q\,S_o\,(8.34)}{A} = \frac{544\,d}{1.085^{(35-T)}\,(t_o)} \qquad \text{6-14}$$

in which

$\frac{QS_o(8.34)}{A}$ = areal BOD loading (lb/acre - day)

A = surface area of the pond (acres)

d = depth of the pond (feet)

T = operating temperature, minimum monthly average (°C)

t_o = optimum detention time for 90 percent BOD removal at 35°C (days)

0.326 = conversion from acre - feet to gallons

8.34 = conversion of gallons to pounds

Equation 6-10 can be rearranged in a similar fashion to result in the same equation as Equation 6-14.

A graphical presentation of Equation 6-14 is illustrated in Figure 6-4. The curves represent the relationship between the areal loading and temperature for 90 percent removal of BOD at three different detention times representing plug flow system, a completely mixed basin and an intermediate case, respectively.

The relationship between the detention time required for 90 percent removal of BOD, the operating temperature, and the influent concentration of BOD is illustrated in Figure 6-5. The detention time of seven days represents a hydraulic flow regime somewhere between plug flow and completely mixed system and was selected for calculation of these data. The curves in Figure 6-4 and 6-5 are applicable over a temperature range some 5°C to 35°C. The effect of temperature on the areal loading which may be applied to ponds and result in a BOD removal of 90 percent is graphically presented in Figure 6-6. Algal growth is generally inhibited in temperatures in excess of 35°C and odors are likely to be produced from the ponds under these conditions. At temperatures below 5°C, an ice cover generally develops during winter conditions and the rate of biodegradation is reduced to a negligible rate. In this case the ponds should be designed to impound the entire winter flow and effluent should be discharged only during the summer and fall months.

The approximate range of effluent BOD concentration is useful for design purposes and the average performance is only of interest over a long period of time. Operating data from various installations were assembled by Aguirre and Gloyna (1971) and are presented graphically in Figure 6-7. The wide scatter in the data represents the influences of temperature, type of wastewater treated, location of ponds, etc. The results of pilot-scale studies of waste stabilization pond performance conducted at the Govalle Wastewater Treatment plant in Austin, Texas are presented in Figure 6-8. These data indicate that the effluent BOD observed at loadings between 30 and 150 lb BOD/acre-day was lower than the maximum effluent BOD reported in the literature and illustrated in Figure 6-7. The data presented in Figure 6-8 indicate that the effluent quality expressed as total BOD is dependent upon the areal loading rate. However, the soluble effluent BOD is independent of the loading within the range of loading conditions used in this study. Two theoretical effluent quality curves for facultative ponds are also included in Figure 6-8. These curves are based on calculated values using Equation 6-14 for operating temperatures of 10°C and 20°C (50°F and 68°F), respectively. The influent BOD (S_o) was assumed to be 200 mg/l and the relationship of optimum detention time, BOD removal and the rate constant. The optimum detention (t_o) was calculated from the relationship presented in Equation 6-15.

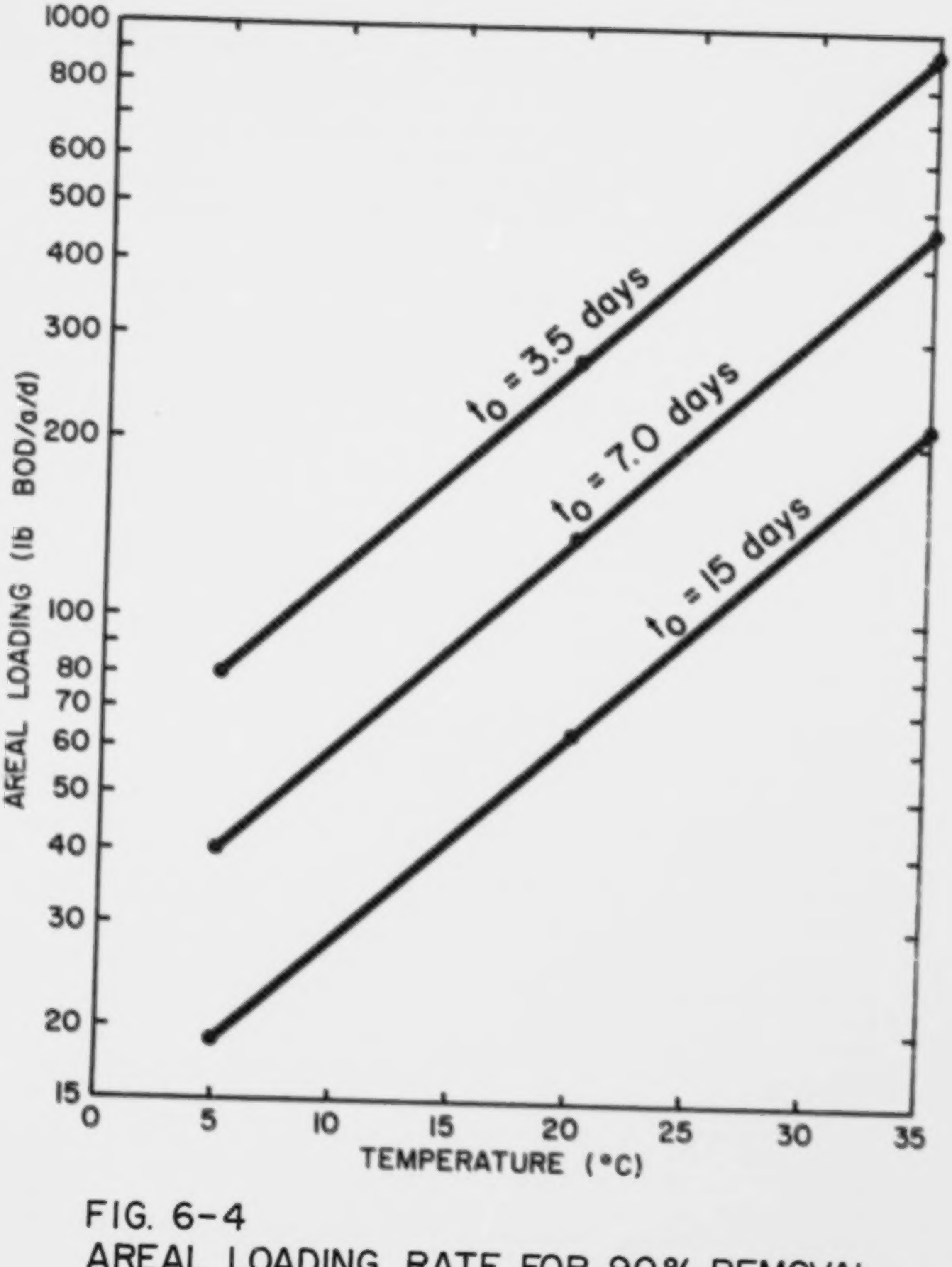

FIG. 6-4
AREAL LOADING RATE FOR 90% REMOVAL

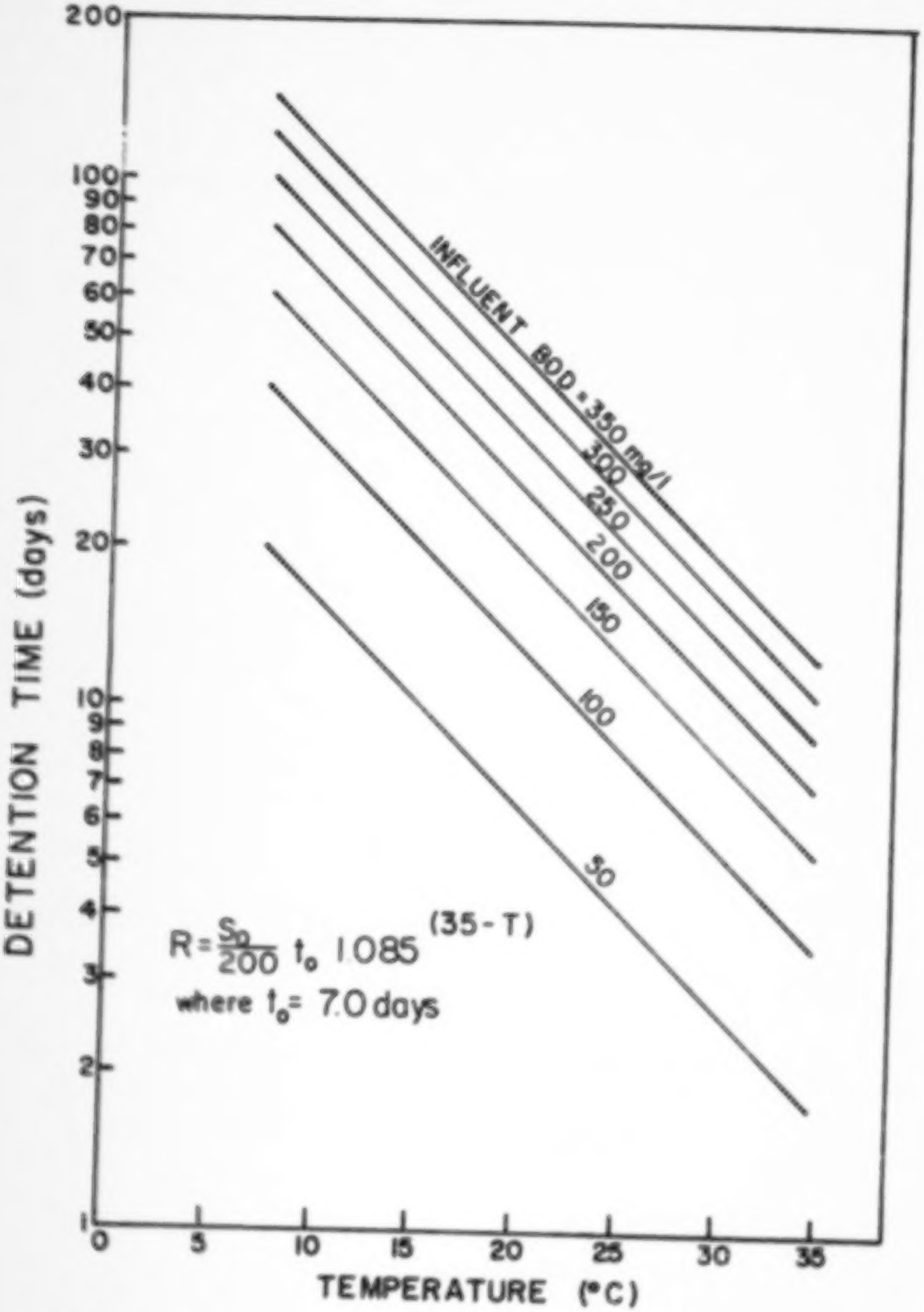

FIG. 6-5
DETENTION TIME FOR 90% BOD REMOVAL

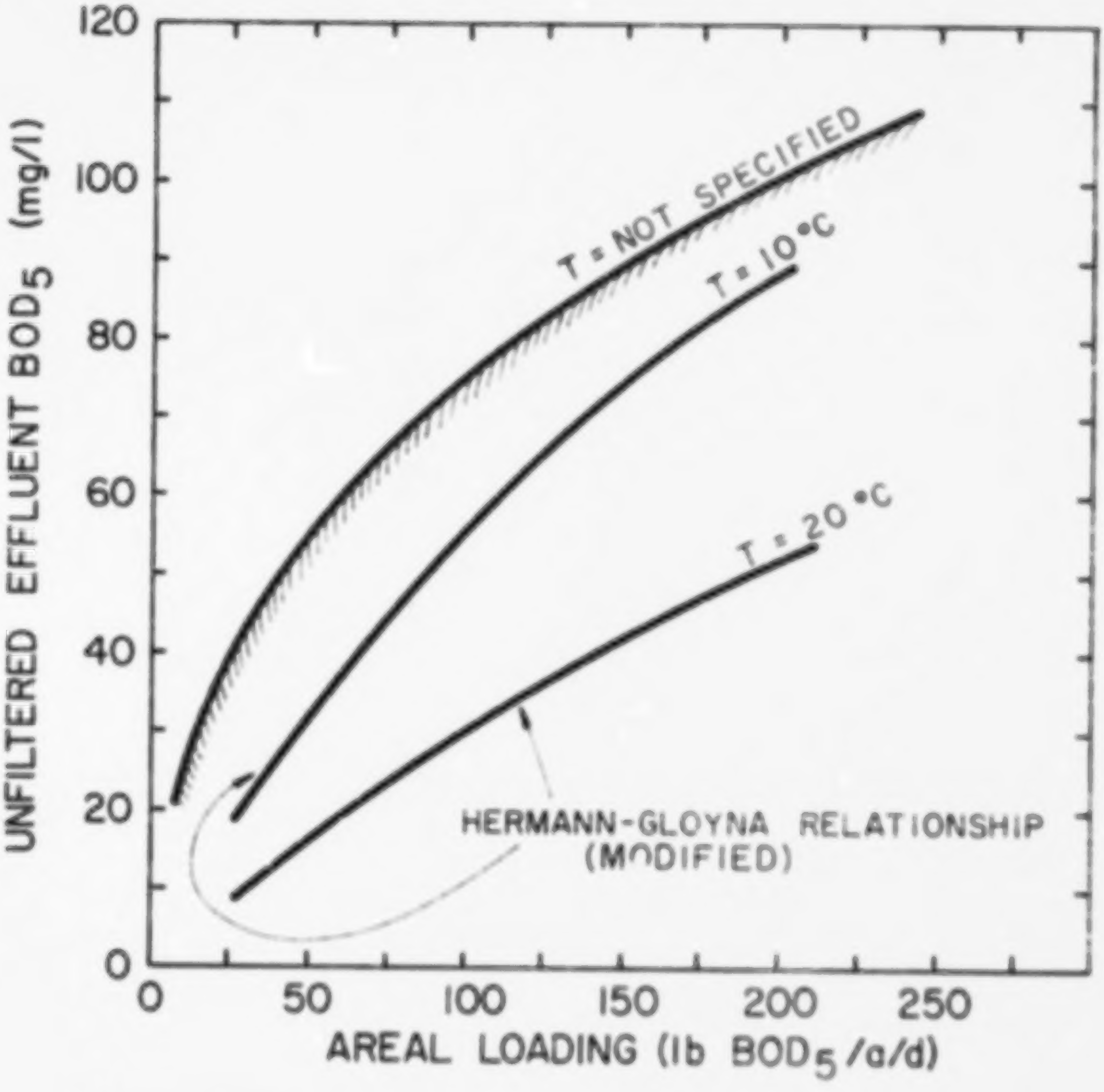

FIG. 6-6
EFFLUENT QUALITY AS A FUNCTION
OF AREAL LOADING

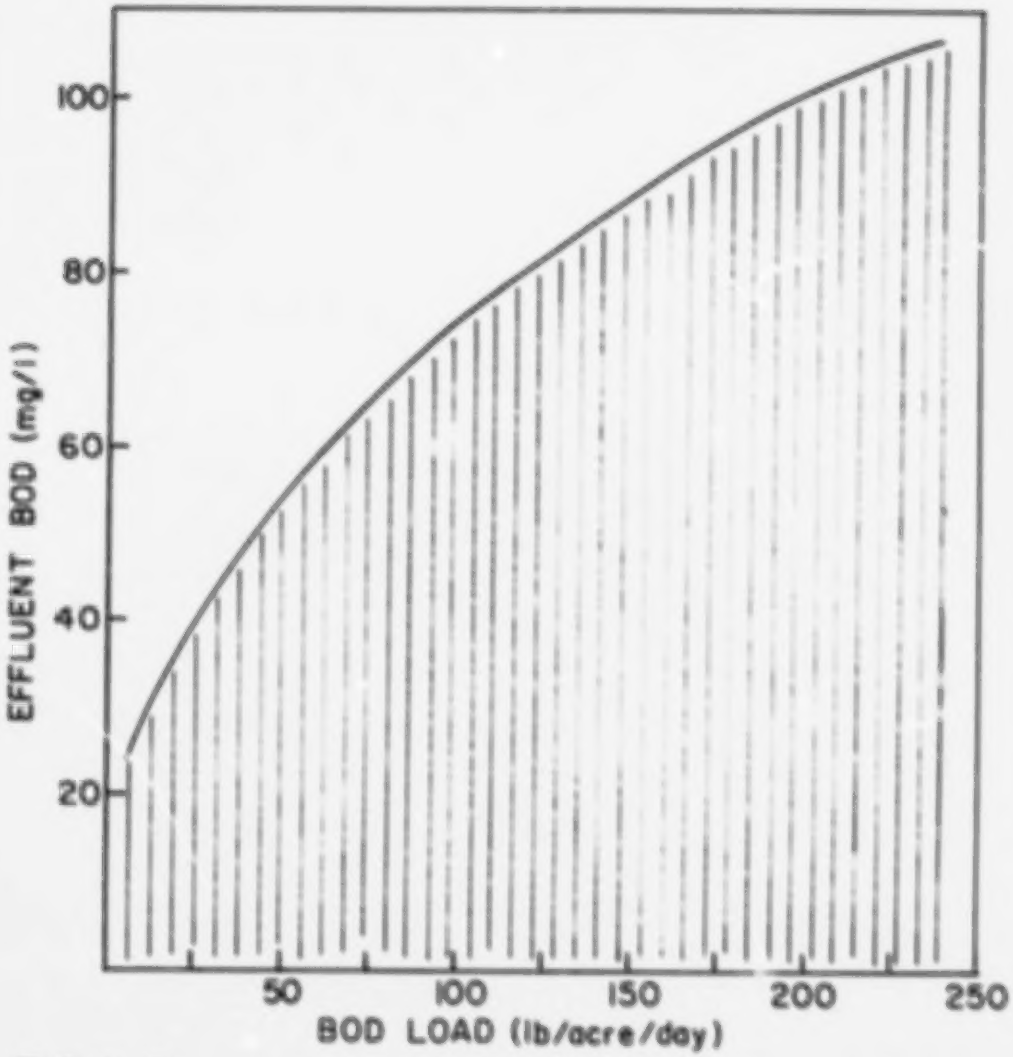

FIG.6-7
RANGE OF LOAD AND EFFLUENT FROM WASTE STABILIZATION PONDS

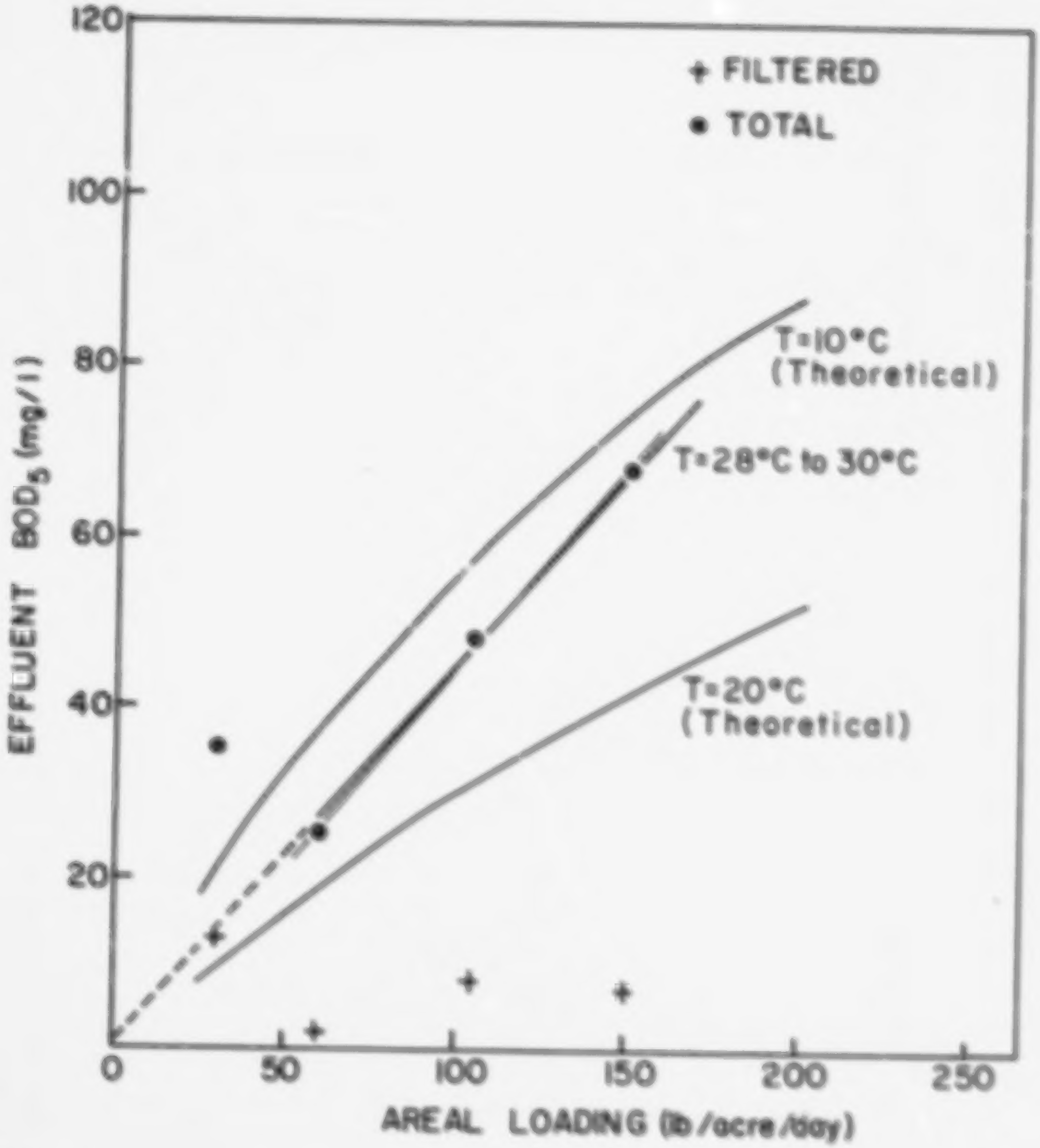

FIG.6-8
EXPERIMENTAL AND THEORETICAL EFFLUENT QUALITY AS A FUNCTION OF AREAL LOADING

$$t_o = \frac{\left(\frac{S_o}{S} - 1\right)}{k_o} \qquad 6\text{-}15$$

The two theoretical curves follow a similar pattern to the other data and indicate that the constants developed are in the useful range. However, for design purposes, these constants should not be over-emphasized since the experimental data observed at pond temperatures of 20°C and above resemble the performance of a pond operating at 10°C. This information indicates that the rate constant in the pilot-scale stabilization pond was much lower than the assumed theoretical value. The calculated rate constants assuming completely mixed conditions are shown in Figure 6-9. The rate constant calculated from the straight line indicates that $k = 0.11$. However, the dashed curve drawn through points indicates that the rate constant might be a function of the loading and is similar to those reported for other biological waste treatment processes.

Design of waste stabilization ponds is generally based on the surface loading rate. In the United States, the recommended surface loading is between 16 and 50 lb BOD/acre-day (Englande, 1969). The lower surface loadings were used primarily in the northern states. However, under favorable environmental conditions, waste stabilization ponds may be effectively operated at much higher loading rates. Oswald (1968) operated ponds in California at surface loadings in excess of 125 lb BOD/acre-day. Horning, et al. (1964) reported that ponds in Ohio could effectively treat municipal wastewater at surface loadings of about 100 lb BOD/acre-day. Canter et al. (1969) & Mills (1961) reported that loadings of up to 200 lb BOD/acre-day have been applied to ponds and produced acceptable effluents in the southern part of the United States. Parker, et al. (1959) recommended loadings of 100 lb of BOD/acre-day and 60 lb BOD/acre-day for summer and winter conditions, respectively in Australia. Ponds of South Africa have been operated successfully at surface loadings of 120 lb BOD/acre-day and as high as 250 lb BOD/acre-day if recirculation was included in the design (Meiring, et al., 1968).

The effluent BOD concentration which can be expected at various loading rates is important in the design of waste stabilization ponds. The maximum effluent BOD based on published data can be related to the loading rate as illustrated in Figure 6-7. The theoretical performance at temperatures of 10°C and 20°C (50°F and 68°F), were used to develop average effluent concentrations and were related to design loading rates as illustrated in Table 6-1.

The soluble effluent BOD of stabilization ponds will range from 5 mg/l to 15 mg/l. However, the effluent suspended solids may range from 40 to 200 mg/l at higher loadings.

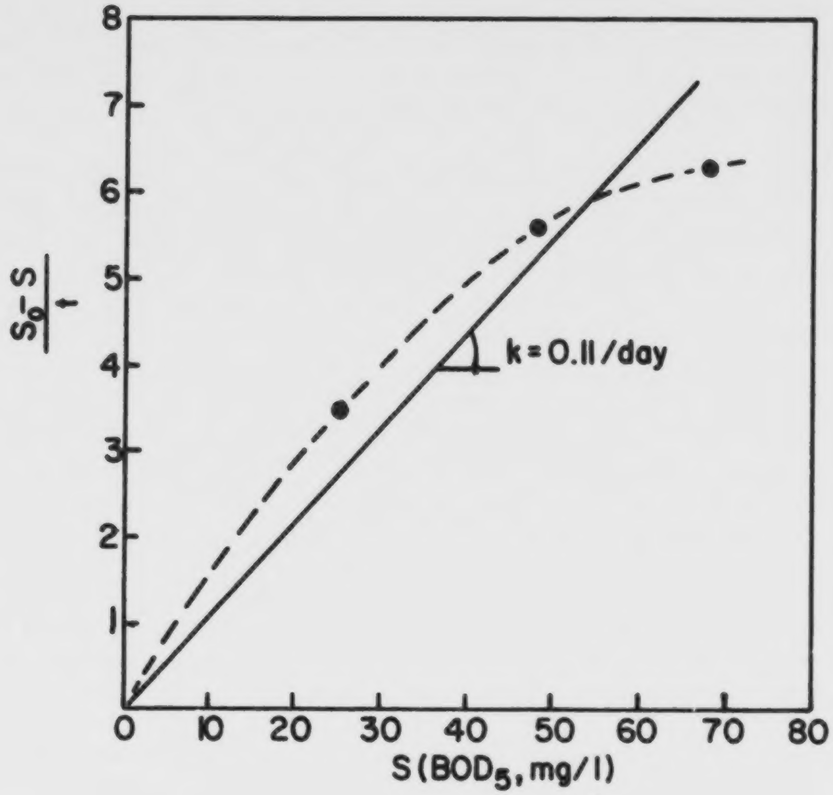

FIG.6-9

CALCULATED BIODEGRADATION RATE CONSTANT, ASSUMING COMPLETELY MIXED CONDITIONS

TABLE 6-1

LOADING RATES AND EFFLUENT BOD FROM SINGLE FACULTATIVE WASTE STABILIZATION PONDS

5-8 FEET DEEP

LOADING RATE lb BOD/acre-day	AVERAGE EFFLUENT BOD Temp. 5-15°C	Temp. 15-25°C	MAXIMUM EFFLUENT BOD
12.5	10-20	10-15	30
25	15-25	10-20	40
50	20-40	10-30	55
100	30-60	20-50	75
150	40-80	25-60	90

An acceptable effluent BOD can be produced by ponds in the southern states at loadings of 50 to 150 lb BOD/acre - day. In the northern states, where the ponds may be operated at low temperatures and under an ice cover during part of the year, loadings of 20 to 50 lb BOD/acre - day should be used.

Sludge accumulation in facultative ponds is a function of the type of wastewater and the quantity of settleable solids entering the ponds, the operating temperature, and the rate of anaerobic degradation of the organic solids in the benthic zone. The rate of synthesis of bacterial cells also affect the solids accumulation in ponds.

MATURATION PONDS

Maturation ponds are usually used as tertiary treatment processes in which the natural self-purification processes result in an improvement of the quality of the effluent. Maturation ponds should not be used to provide additional treatment to the effluent of under-designed conventional biological processes. In maturation ponds, the concentration of bacteria, viruses and nutrients, as well as suspended solids, can be reduced and thereby improve the quality plant effluents.

The performance of maturation ponds is expressed primarily in terms of reduction in the number of bacteria. The bacterial die-off rate is affected by the detention times, sunlight, bactericidal agents released by algae, temperature, dissolved oxygen, pH, predation, nutrient depletion, and toxic materials (Davis and Gloyna, 1970). Bacterial removal of between 90 and 99.9 percent have been reported in the literature. The total coliform density of municipal wastewater expressed as the Most Probable Number (MPN) is usually in a range of 10^6 to 10^8 per hundred milliliters. In most cases the standards for discharge to receiving streams are set at an MPN of 10^6 per hundred milliliters. However, to meet these high effluent standards, the reduction in the concentration of coliform organisms would have to be about 10^5 per 100 ml, or approximately 99.999 percent. These high efficiencies cannot be guaranteed by a simple maturation pond even at an extremely long detention time. In general, it is agreed that the detention time for maturation ponds is between ten to 15 days.

PONDS IN SERIES

Waste stabilization ponds can be operated in series to minimize the possibility of short circuiting and the discharge of partially treated effluent from stabilization ponds. In general, the effluent of the last pond of a series is of better quality than that from a single pond which had the identical detention time as the ponds in series. The results of pilot-scale investigation at the Govalle Wastewater Treatment Plant in which a combination of ponds were operated in series and in parallel indicate that the system consisting of three ponds produced an effluent quality which is better than that resulting from two ponds in series. A schematic diagram of the pilot-scale stabilization pond systems are presented in Figure 6-10. One of the systems included an anaerobic pond, a facultative pond, and a maturation pond in series. The second system included a facultative pond which was deeper at the center to provide an anaerobic zone of the same volume as the anaerobic pond in system one, followed by a maturation pond. The third system included a facultative pond and a maturation pond. The results of these investigations indicate that the three pond system produced the best effluent. Approximately 50 percent of the BOD and organic carbon in the influent wastewater was removed in the anaerobic pond. Therefore, the BOD load to the facultative pond was reduced and the quantity of algal cells produced in the facultative pond was also reduced, resulting in an effluent of better quality. It might be advisable to install two anaerobic ponds in order to provide sufficient time for the sludge accumulated in one pond to undergo more complete anaerobic degradation and permit disposal of the sludge on the land.

DESIGN FACTORS

The design of waste stabilization ponds for the treatment of municipal wastewaters is based essentially upon the total quantity of wastewater to be treated and in the influent BOD concentration. It is also essential that any industrial wastes or toxic materials present in the incoming wastewater be identified to minimize potential toxic effects on the algae. The basic design information required includes:

(a) total wastewater flow per day (MGD)
(b) concentration of BOD (mg/l)
(c) ratio of five-day BOD to ultimate BOD
(d) the minimum and maximum air temperatures
(e) if freezing occurs, the duration of time during which an ice cover can be expected
(f) the presence of any potentially toxic material

The design of waste stabilization pond systems is affected by the effluent standards as well as the stream standards established by regulatory agencies. The design of waste stabilization ponds also is dependent upon the availability and location of suitable land which can be used as a plant site.

Waste stabilization pond systems are applicable for the treatment of municipal wastewaters in those areas in which effluent standards have

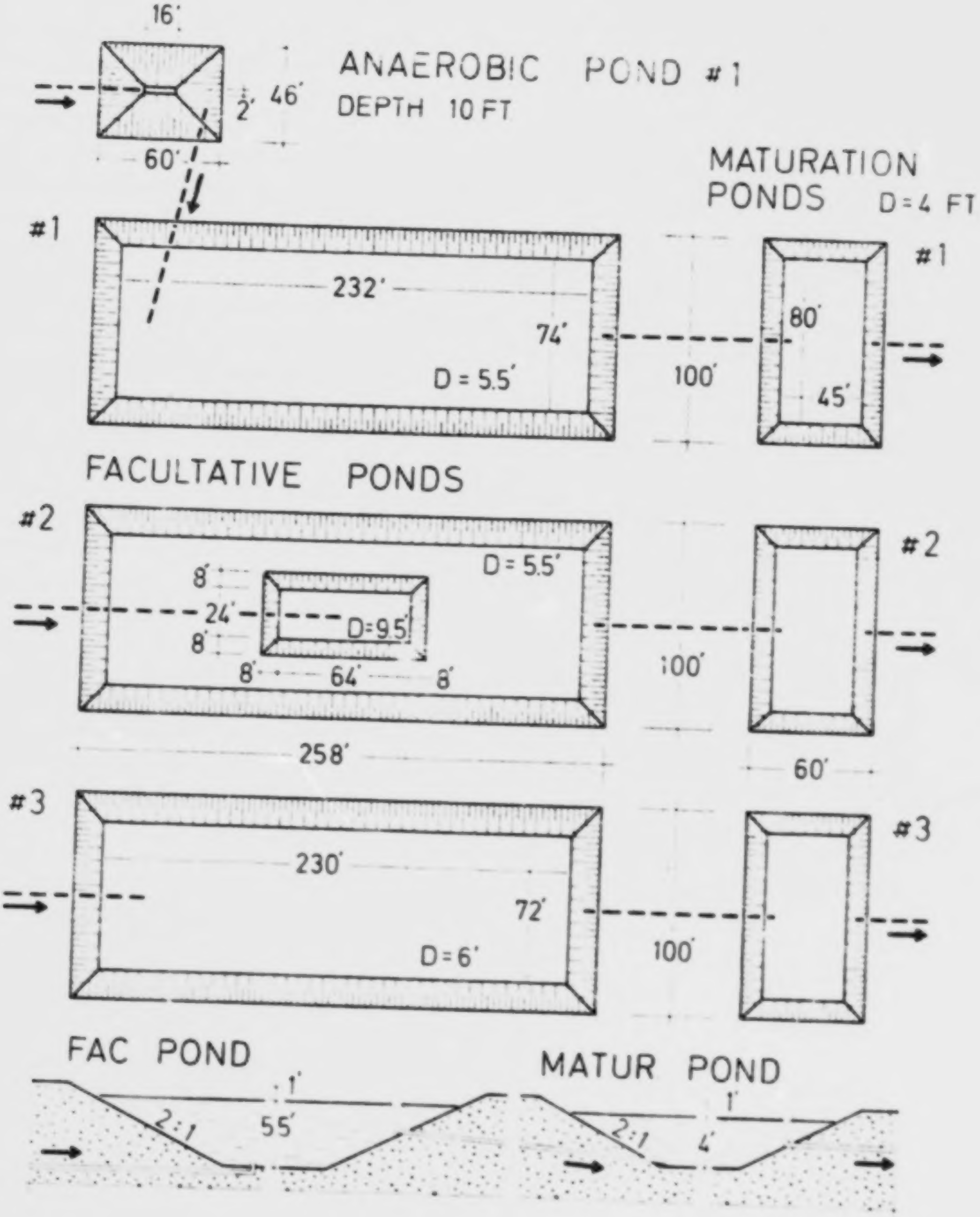
16'
ANAEROBIC POND #1
DEPTH 10 FT
2'
46'
60'
MATURATION
PONDS D=4 FT
#1
232'
74'
D = 5.5'
100'
80'
45'
#1
FACULTATIVE PONDS
#2
8'
24'
8'
D=9.5'
8'
64'
8'
D = 5.5'
100'
#2
258'
60'
#3
230'
72'
D = 6'
100'
#3
FAC POND
1'
2:1
5.5'
MATUR POND
1'
2:1
4'

FIG.6-10

PILOT WASTE STABILIZATION PONDS

been established which will permit an effluent BOD and suspended solids (algal cells) concentration of more than 20 mg/l. The following design procedure is proposed:

(a) The regulatory agencies accept the fact that reliable effluent quality can be produced by waste stabilization ponds, but with the understanding that the effluent quality will vary between summer and winter operation. The pond system be designed to meet the effluent standards as closely as possible.

(b) Relatively inexpensive land should be available for the extension of the plant as population increases.

(c) The required volume for facultative ponds can be calculated using Equation 6-9 and the areal loading can be calculated from Equation 6-14. The curves presented in Figures 6-4, 6-5, and 6-7 can be used to size the particular ponds.

(d) A maturation pond should follow a facultative pond to improve the bacteriological quality of the plant effluent.

DESIGN EXAMPLE

Two waste stabilization pond systems are designed to treat municipal wastewater. This example illustrates the difference between the performance of a single facultative pond followed by a maturation pond and a system including an anaerobic pond followed by a facultative pond and a maturation pond. The characteristics of the wastewater are:

Influent wastewater flow = 1 MGD
Influent concentration of BOD_5 = 200 mg/l
Influent concentration of ultimate BOD = 300 mg/l
Influent concentration of suspended solids = 200 mg/l

The following climatological data are used in the design of the pond systems:

Average monthly temperature range = 10 to 30°C
Average temperature of the coldest month = 10°C
The rainfall is assumed to equal the evaporation

Two Pond System

The design procedure for the facultative pond is as follows:

The required detention time for the facultative pond can be derived from Figure 6-5. At a temperature of 10°C and an influent BOD of 300 mg/l, the required detention time is 82 days.

The pond volume required is $V = Q\,t$

$$V = 1 \times 10^6 \ \frac{\text{gal}}{\text{day}} \left(\frac{\text{cu ft}}{7.48 \text{ gal}}\right) 82 \text{ day} = 11 \times 10^6 \text{ cu ft}$$

The surface area of the pond assuming a five-foot operating depth is

$$A_{\text{sur}} = \frac{V}{d} = \frac{11 \times 10^6 \text{ cu ft}}{5 \text{ ft.}} = 2.2 \times 10^6 \text{ sq ft}$$

or

$$2.2 \times 10^6 \ \frac{\text{sq ft (acre)}}{43{,}560 \text{ sq ft}} = 50.5 \text{ acres}$$

One foot of depth is added for sludge storage; therefore, the volume is:

$$V = 6 \text{ ft } (2.2 \times 10^6 \text{ sq ft}) \ \frac{7.48 \text{ gal}}{\text{cu ft}} = 99 \times 10^6 \text{ gal}$$

The organic load to the pond is:

$$\frac{1 \times 10^6 \text{ gal}}{10^6 \text{ day}} \left(\frac{300 \text{ mg}}{\text{L}}\right) \left(\frac{8.34 \text{ lb}}{\text{gal}}\right) = \frac{2500 \text{ lb BOD}_u}{\text{day}}$$

The areal load is therefore:

$$\frac{2500 \text{ lb BOD}_u / \text{day}}{50.5 \text{ acres}} = \frac{49.5 \text{ lb BOD}_u}{\text{acre-day}}$$

The volumetric load is:

$$\frac{2500 \text{ lb BOD}_u / \text{day}}{13.2 \times 10^6 \text{ cu ft}} = \frac{0.19 \text{ lb BOD}_u}{100 \text{ cu ft - day}}$$

Three Pond System

The required detention time of the anaerobic pond to minimize odor problems is:

$$t = 5 \text{ days}$$

The volume of the anaerobic pond must include sludge storage; therefore:

Total Volume = Liquid Volume + Sludge Storage

$$\text{Liquid Volume} = 5 \text{ days } \left(1 \times 10^6 \ \frac{\text{gal}}{\text{day}}\right)\left(\frac{\text{cu ft}}{7.48 \text{ gal}}\right) = 0.67 \times 10^6 \text{cu ft}$$

Sludge storage volume is based on the assumption that 100 percent of the solids are removed and concentrate to three percent solids. The required detention time of the solids for anaerobic degradation at 10°C is 90 days.

Dry solids removed:

$$200 \text{ mg/l } \left(\frac{1 \times 10^6}{10^6} \ \frac{\text{gal}}{\text{day}}\right) \ \frac{8.34 \text{ lb}}{\text{gal}} = 1668 \ \frac{\text{lb}}{\text{day}} \ \text{solids}$$

The volume of sludge is:

$$1668 \ \frac{\text{lb solids}}{\text{day}} \left(\frac{100 \text{ lb sludge}}{3 \text{ lb solids}}\right)\left(\frac{\text{cu ft}}{62.4 \text{ lb sludge}}\right) 90 \text{ days} = 0.8 \text{x} 10^5 \text{cu}$$

Therefore, the Total Volume =

$$(7 \times 10^5) + (0.8 \times 10^5) = 7.8 \times 10^5 \text{ cu ft, or } 5.85 \times 10^6 \text{ gal}$$

$$\text{The total hydraulic detention time} = \frac{5.85 \times 10^6 \text{ gal}}{1 \times 10^6 \text{ gal/day}} = 5.85 \text{ days}$$

A depth of ten feet in anaerobic pond results in a surface area of:

$$A_{sur} = \frac{0.78 \times 10^6 \text{ cu ft}}{10 \text{ ft}} \quad \frac{\text{acre}}{43{,}560 \text{ sq ft}} = 1.79 \text{ acres}$$

The loading to the anaerobic pond is:

$$\text{Areal loading} = \frac{2500 \text{ lb BOD}_u\text{/day}}{1.79 \text{ acres}} = \frac{1395 \text{ lb BOD}_u}{\text{acre - day}}$$

$$\text{Volume loading} = \frac{2500 \text{ lb BOD}_u\text{/day}}{0.78 \times 10^6 \text{ cu ft}} = \frac{3.2 \text{ lb BOD}_u}{1000 \text{ cu ft - day}}$$

The concentration of BOD entering the facultative pond assuming 50 percent BOD_u removal in the anaerobic pond is:

$$\text{BOD}_u = 0.5 \ (300 \text{ mg/l}) = 150 \text{ mg/l}$$

The required detention time from Figure 6-5 for an operating temperature of 10°C and $\text{BOD}_u = 150$ mg/l is:

$t = 41$ days

The required volume therefore is:

$$V = (41 \text{ days}) 1 \times 10^6 \frac{\text{gal}}{\text{day}} \; \frac{\text{cu ft}}{7.48 \text{ gal}} = 5.48 \times 10^6 \text{ cu ft}$$

At a pond depth of five feet, the surface area is:

$$A_{sur} = \frac{5.48 \times 10^6 \text{ cu ft}}{5 \text{ ft}} \; \frac{\text{acre}}{43{,}560 \text{ sq ft}} = 25 \text{ acres}$$

Provision for sludge storage in the facultative pond following an anaerobic pond is not necessary.

The organic load to the facultative pond is:

$$BOD_u = \frac{1 \times 10^6 \text{ gal}}{10^6 \text{ day}} \left(\frac{150 \text{ mg}}{1}\right) \left(8.34 \frac{\text{lb}}{\text{gal}}\right) = 1250 \text{ lb/day}$$

The loadings to the facultative pond are:

$$\text{Areal loading} = \frac{1250 \text{ lb } BOD_u}{25 \text{ acre - day}} = 50 \frac{\text{lb } BOD_u}{\text{acre-day}}$$

$$\text{Volume loading} = \frac{1250 \text{ lb } BOD_u}{5.48 \times 10^6 \text{ cu ft - day}} = \frac{0.28 \text{ lb } BOD_u}{1000 \text{ cu ft-day}}$$

The comparison of the two systems is presented in Table 6-2. The area requirements for the anaerobic-facultative-maturation pond system is almost 50 percent less than that required for the facultative-maturation-pond system.

TABLE 6-2

COMPARISON OF DESIGN REQUIREMENTS FOR A CONVENTIONAL FACULTATIVE POND AND AN ANAEROBIC-FACULTATIVE POND SYSTEM

SYSTEM	DETENTION TIME (days)	VOLUME (MG)	SURFACE AREA (acres)	AREAL LOADING $\left(\frac{\text{lb BOD}_u}{\text{acre-day}}\right)$	VOLUMETRIC LOADING $\frac{\text{lb BOD}_u}{\text{1000 cu ft - day}}$
Facultative	82	99	50.5	49.5	0.19
Maturation	15	15	12	----	----
Total	97	114	62.5		
Anaerobic	5.85	5.85	1.79	1395	3.2
Facultative	41	41	25	50	0.28
Maturation	15	15	12	----	----
Total	61.85	61.85	38.79		

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SYMBOLS AND ABBREVIATIONS

a	fraction of substrate converted to bacterial mass
a'	oxygen consumed per unit of substrate removed
b	substrate removed per unit of time, by endogenous respiration
b'	endogenous respiration rate
BOD	biochemical oxygen demand
BODF	filtered (assumed to be soluble) BOD_5
BODT	total BOD_5
^{o}C	degrees centigrade
c	oxygen concentration
COD	chemical oxygen demand
CODF	filtered (assumption: soluble) COD
CODT	total COD
^{o}F	degrees Fahrenheit
IC	inorganic carbon
k	reaction rate coefficient, related to MLSS
K	reaction rate coefficient, MLSS included
k_a	reaction rate coefficient, related to active MLSS
K_a	reaction rate coefficient, active MLSS included
K_h	heat transfer coefficient
k_m	substrate-concentration at which the reaction-rate equals one-half k_{max}
k_R	k_{max} divided by a
k_v	reaction rate coefficient, related to MLVSS

k_{max}	maximum removal rate, Monod - Michaelis - Menten equation
K_v	reaction rate coefficient, MLVSS included
K_v	equals K, K_a
K_T	reaction rate coefficient, at temperature T, °C
K_{20}	reaction rate coefficient, at 20°C
MLSS	mixed liquor suspended solids
MLVSS	mixed liquor volatile suspended solids
NH_3	ammonia
NO_{2+3}	nitrite + nitrate
OUR	oxygen uptake rate
Q	flow
R	oxygen used, per unit weight of mixed liquor and per unit of time
S	soluble substrate concentration, tank
S_{o_s}	soluble substrate concentration, influent
ST_e	total substrate concentration, supernatant after settling
ST_o	total substrate concentration, influent
ΔS	soluble substrate removed, S_o - S
SS	suspended solids
t	hydraulic detention time, t = V/Q
T	temperature
TKN	totak Kjeldahl nitrogen
TC	total carbon
TOC	total organic carbon

TOCF	filtered (assumption: soluble) TOC
TOCT	total TOC
TPO_4	total phosphate
t_v	variable time
X	MLSS
X_a	active MLSS
X_o	SS in influent wastewater
X_{oa}	active SS in raw wastewater
X_v	volatile MLSS
V	volume of reactor
θ	temperature-correction-coefficient

When other symbols and abbreviations are used in the text, they are explained.